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Chemically Active Fluid-Bed Process for Sulphur Removal During Gasification of Heavy Fuel Oil - Second Phase

Esso Research Centre, Abingdon (England)

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16. Abstract: The report describes the second phase of studies on the CAFB process for desulfurizing gasification of heavy fuel oil in a bed of hot lime. The first continuous pilot plant test with U.S. limestone BCR 1691 experienced local stone sintering and severe production of sticky dust during startup. Batch tests confirmed that BCR 1691 produced more dust than the purer Denbighshire or U.S. BCR 1359 stones. With BCR 1691, 10 times more dust was produced during kerosene combustion at 870°C than during gasification/regeneration. The continuous pilot plant was modified to improve operability under dusty conditions: 132 gasification hours were spent in a second run with Denbighshire and BCR 1691 stones in six operating periods, the longest being 109 hours. Sulfur removal efficiency was comparable for the two stones, ranging from 60 to 95%. Regenerator performance was less satisfactory than in earlier tests. A poor sulfur material balance indicates need for improved analytical procedures. Total CAFB development through a large demonstration test will probably take about 6-7 years and require $3,320,000 in engineering effort.

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CHEMICALLY ACTIVE FLUID-BED PROCESS FOR SULPHUR REMOVAL DURING GASIFICATION OF HEAVY FUEL OIL SECOND PHASE

by

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ABSTRACT

This report describes the second phase of studies on the CAFB process for desulphurising gasification of heavy fuel oil in a bed of hot lime.

The first test of the continuous pilot plant with U.S. limestone BCR 1691 was hampered by local stone sintering and severe production of a sticky dust during start up conditions. Batch tests confirmed that BCR 1691 produced more dust than either of the higher purity Denbighshire or U.S. BCR 1359 stones. With BCR 1691 dust production rate was tenfold higher during kerosene combustion at 870 deg. C than during gasification/regeneration cycles.

Modifications were made to the continuous pilot plant to improve operability under dusty conditions, and a total of 112 gasification hours was made in a second run with Denbighshire and BCR 1691 stones in six operating periods, the longest being 109 hours.

Sulphur removal efficiency was comparable for the two stones, ranging from 60 to over 95%. Regenerator performance was less satisfactory than in earlier tests, and a poor sulphur material balance indicates need for improved analytical procedures.

An engineering scoping study estimate that total CAFB development through a large demonstration test will take about 6-7 years and require $1,120,000 in engineering effort.

This report was submitted as a requirement of Contract No. 68-02-0(FA) by Esso Research Centre, England under the sponsorship of the Environmental Protection Agency. Work during the first half of the above contract is included. This was completed in April 1973.

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Finally, they acknowledge the assistance given by other Esso Petroleum Company personnel in operating the pilot plant, in maintenance of equipment, and in chemical analysis of test samples.

SECTION 1

CONCLUSIONS

TASK I

1. Stability and quality of continuous pilot plant operations were greatly improved by modifications which included uninterrupted limestone addition, regenerator back pressure control, flue gas recycle scrubbing, nitrogen quench for regenerator over temperature protection, and a positive pressure pilot flame for the main burner.

2. No significant difference was found between the sulphur removal ability of BCR 1691 and Denbighshire limes at comparable operating conditions in the continuous pilot plant.

3. Lime replacement rate has a significant effect on sulphur removal efficiency with stoichiometric Ca/S replacement giving about 80% sulphur removal at 18 to 25 inch water gauge (w.g.) bed depth.

4. Increasing gasifier bed depth increases the rate of lime loss into the boiler and thus may obscure any effect of depth on sulphur absorption. No correlation of sulphur removal efficiency with bed depth was found in the region of 18 to 25 inches w.g.

5. Regenerator performance in the continuous pilot plant was not as good as experienced using Denbighshire stone in Phase I studies. Quality of fluidisation was poorer, and selectivity of oxidation of CaS to CaO plus SO_2 was lower. Regenerator off gas SO_2 concentrations remained below 1% for much of the test. Also, less than half the sulphur removed from the fuel (as measured by flue gas analysis) was recovered as SO_2 by the regenerator (as indicated by regenerator gas analysis). The reduction in regenerator performance may be due to the absence of preferential fines circulation from gasifier cyclone to regenerator in this test.

6. Regenerator SO_2 output as measured by gas analysis was significantly lower than values calculated from solids analysis.

7. In the continuous pilot plant test, metals removal from the fuel by the lime averaged 16% for sodium, 75% for nickel, and essentially 100% for vanadium.

8. Cyclone fines return performance deteriorated during the third day of the second continuous pilot plant test in this phase of studies, and subsequent periods experienced high loss rates of lime from the gasifier into and through the boiler.

9. In the continuous pilot plant with Denbighshire lime, most of the gasifier solids losses were recovered in the boiler or in the external flue gas cyclone. With BCB 1691, most of the time losses passed through both boiler and external cyclone.

10. Variations in cyclone fines return system performance during the second continuous pilot plant run were reflected by changes in bed particle size distribution.

11. Stainless steel cyclone liners are not satisfactory as a means of providing a smooth cyclone surface. In the pilot plant test they failed under decoking conditions and provided a surface for increased carbon deposition.

12. Retention of fine lime within the gasifier and avoidance of deterioration in cyclone performance are the major operating problems remaining in the CAFB development.

13. Operation of the gasifier bed at high enough levels to assure a high concentration of solids in the gas entering the cyclones reduces the rate of carbon deposition in these ducts and increases the operating interval between decoking operations.

14. Combustion of CAFB gasifier product in the pilot plant burner produces less nitrogen oxides (165 ppm average) than direct combustion of the fuel oil (261 ppm average).

15. The heat release by fuel partial combustion in the gasifier is approximately 1130 Btu/lb at 25% of stoichiometric air. This value agrees with thermodynamic estimates based on the fractions of carbon and hydrogen oxidised and the amount of CO produced.

16. Sintering of BCB 1691 stone during unit start-up is possible but can be avoided by adjusting the gasifier/gas generator pressure balance to avoid direct impingement of the start-up burner flame into solids transfer ducts.

PHASE II

1. Lime from BCB 1691 stone produces much more dust under comparable CAFB fluidisation conditions than either of the higher purity limes BCB 1359 and Denbighshire.

2. Dust production with BCB 1691 lime is particularly severe during combustion with kerosene at 870 deg. C, the normal CAFB start-up condition.

3. Kerosene combustion at 1050 deg. C causes less dust production than combustion at 870 deg. C both with BCB 1691 and with Denbighshire lime.

4. The fine dust produced during 870 deg. C combustion with MCR 1691 lime is sticky in nature and clings to pipe walls and cyclone internals unless mechanical force such as rapping is employed to dislodge it. Under gasification conditions the dust is not sticky, and is produced at a lower rate.

5. Sulphur removal activity of MCR 1359 appears to be intermediate between Denbighshire and MCR 1691 limestones.

PHASE III

An engineering scoping study by Esso Engineering indicates that total PAFB development through a 100 + MW demonstration test period is expected to take about 6½ years and require $1,320,000 in engineering effort. Optimistically the development time might be reduced to 4½ years with a cost of $1,520,000, but risks associated with the large unit would be correspondingly increased.

SECTION II

RECOMMENDATIONS

1. Pilot plant efforts should be shifted from BCR test stone to BCR 1359 which poses much less severe operating problems from a dust production and potential sintering stand point.

2. Future tests of potential CAFB limestones should include measurements of their dust producing characteristics under the fully combusting conditions encountered during startup and hot standby operations.

3. The pilot plant should be modified to permit reexamination of the effect of preferential circulation of gasifier fines direct to the regenerator.

4. Any CAFB installation should provide steady addition of limestone to prevent temperature variations caused by the changing heat load.

5. An emergency regenerator quench system should be included in CAFB installations to prevent sintering and agglomeration by temperature upsets.

6. A scrubber is needed in CAFB installations for the flue gas recycle to prevent fouling the recycle system, blowers, and air distributors.

7. Means of ensuring a high circulation rate of solids through the gasifier cyclones should be provided to decrease carbon deposition rate and to increase run length between decoking steps.

8. Investigation of a means to provide smooth cyclone interior surfaces should be continued.

9. Boiler gas and regenerator gas rate measurements, sampling systems, and analytical procedures should undergo continuing study and revision until discrepancies in sulphur material balance values are resolved. In particular, a high velocity, hot flue gas sample system with hot cyclone and filter is recommended for the flue gas analysis procedure.

10. Regenerator operation should be tested at lower air rates to confirm if reduced CaO conversion level will improve selectivity of CaS oxidation to CaO and reduce the quantity of $CaSO_4$ returned to the gasifier.

11. An increased level of effort on data analysis and engineering study of various features of the process is recommended to improve understanding of results already obtained.

SECTION II:

INTRODUCTION

GENERAL

The Chemically Active Fluid Bed process is a means of avoiding sulphur oxide pollution while using heavy fuel oil for production of power. The process uses a fluidised bed of lime particles to convert the oil into a hot, low sulphur gas ready for combustion in an adjacent boiler. Sulphur from the fuel is absorbed by the lime which can be regenerated for reuse. During lime regeneration the sulphur is liberated as a concentrated stream of SO_2 which may be converted to acid or elemental sulphur.

Exploratory work on the CAFB began at the Esso Research Centre, Abingdon (ERCA) in 1966. Batch reactor fuel and limestone screening studies, a variable study with U.K. limestone BCR 191, and initial operation of a pilot plant incorporating continuous gasification and regeneration were conducted from 1970 to 1972 under Office of Air Programmes (OAP) contract 70-45 and were described in the final report [1] for that contract, dated June 1972.

The current contract represents the second phase of work on this project. This interim report covers work on this phase for the period July 1, 1972 through April 1973.

GASIFIER CHEMISTRY

When heavy fuel oil is injected into a bed of fluidised lime under reducing conditions at about 900 deg. C, it vaporises, cracks, and forms a series of compounds ranging from H_2 and CH_4 through heavy hydrocarbons to coke. The sulphur contained in the oil forms compounds such as H_2S, COS and CS_2 with H_2S predominating. The sulphur compounds react with CaO to form CaS and gaseous oxides.

For example:

$$CaO + H_2S \longrightarrow CaS + H_2O$$

The equilibrium [1] for this reaction is far to the right. With a fuel containing 4% sulphur the equilibrium permits a desulphurising efficiency greater than 90% up to 1100 deg. C. Other factors however limit gasification temperature to the range of about 850 to 900 deg. C. where the equilibrium sulphur removal would be about 99%.

In the shallow fluidised bed of the gasifier there is a rapid circulation of lime between top and bottom. Indications are that coke is laid down on the lime in the upper portion of the fluid bed by oil cracking and coking reactions and that this coke burns off in the lower portion where oxygen is supplied by the air distributor.

Gasification conditions of temperature and air-fuel ratio must be chosen to maintain a balance between the rate of coke and carbon deposition and the rate of carbon burnoff. Broadly, this balance is met at gasification temperatures in the range of 850 to 900 deg. C. and air-fuel ratios around 20% of stoichiometric. Lower air fuel ratios are operable at the upper end of the temperature range, and higher air-fuel ratios are needed as temperature is reduced.

Much of the oxygen entering the gasifier is consumed in oxidising coke to CO and CO_2 near the distributor. Of course, some enters other regions of the bed where it reacts with H_2 and hydrocarbons to form water and more carbon oxides. The final product from the gasifier is a hot combustible gas containing N_2, hydrocarbons CH, CO, H_2O, and H_2. Most of the energy released by partial combustion of the fuel is retained by this gas as sensible heat.

Only a portion of the CaO in the lime is reacted on each pass of solids through the gasifier. Good sulphur absorption reactivity has been obtained with up to 20% of calcium reacted in single cycle batch reactor tests, but in the continuous unit, the average extent of calcium conversion to sulphide is held to less than 10%.

When a single batch of lime is cycled between gasification and regeneration conditions it gradually loses activity for sulphur absorption. The activity of the bed can be maintained if some of the lime is purged each cycle and replaced by fresh material. Reactivity of the bed is therefore a function of the lime replacement rate. The replacement lime is usually added to the gasifier as limestone which calcines in situ.

Vanadium from the fuel oil deposits on the lime during gasification. Experimental evidence is that practically all of the fuel vanadium can remain fixed with the lime.

REGENERATOR CHEMISTRY

Calcium sulphide is regenerated to calcium oxide by air oxidation:

$$CaS + 3/2\ O_2 \longrightarrow CaO + SO_2$$

$$\Delta H = -109.5\ K\ cal/mole$$

A competing reaction also consumes oxygen and forms calcium sulphate:

$$CaS + 2\ O_2 \longrightarrow CaSO_4$$

$$\Delta H = -220.2\ K\ cal/mole$$

Both reactions are strongly exothermic. A third reaction between the solid species is also possible:

$$CaS + 3CaSO_4 \longrightarrow 4CaO + 4SO_2$$

$$\Delta H = 221.5\ K\ cal/mole$$

This reaction is strongly endothermic.

The equilibrium constants (Appendix A, Reference 1) for these reactions determine the maximum partial pressure of SO_2 which can exist in equilibrium with mixtures of CaS, CaO, and $CaSO_4$ at any given temperature. These equilibria also determine a relationship between regenerator temperature and the maximum theoretical selectivity of oxidation of CaS to CaO.

At low oxidation temperature, the equilibrium SO_2 partial pressure is too low to permit all the oxygen supplied to leave in the form of SO_2. The excess oxygen then goes to for $CaSO_4$. Experimental oxidation selectivities are lower than the theoretical maximum, probably because of contacting and kinetic factors.

Since each sulphided lime particle passes through a range of temperatures and oxygen concentrations during its transit through the regenerator, it is exposed, on average, to less favourable selectivity conditions than those at the top of the bed.

Calcium sulphide oxidation selectivities to CaO of 70 to 80% and regenerator SO_2 concentrations of 8 to 10% have been achieved in pilot plant operations at regenerator temperatures in the range of 1040 to 1070 deg. C.

During the conversion of CaS to CaO and $CaSO_4$ there is evidence for existence of a transient liquid state [2]. If air is passed through a hot static bed containing CaS, some of the particles will agglomerate into lumps during the regeneration reaction. Agglomeration does not occur if the bed is vigorously fluidised.

PREVIOUS EXPERIMENTAL WORK

A basis for the CFB process had been established by experiments in 7-inch I.D. batch reactors at CRCA prior to 1970.

During 1970, two new batch reactor units were constructed for the CAF contract. Work in these batch units established the suitability for CFB of a Venezuelan fuel oil available in the U.S., and selected the better of two U.S. limestones suggested by CAF. Both of the U.S. stones, BCR 1690 and BCR 1691, were lower in CaO content than the U.K. stones tested previously. In cycle tests the BCR 1691 stone gave sulphur removal activity comparable to that of the higher purity U.K. stones at equal CaO/S ratios. The BCR 1690 stone was found to be unsuitable in three respects. It gave lower sulphur recovery at equal CaO ratios; it attrited badly and it sintered and agglomerated during regenerator cycles. The BCR 1691 stone therefore was selected for further study. During 1971 an extensive study of fluidization variables was conducted in the batch reactors with this stone. Tests with fresh beds screened the effects of major variables including air/fuel ratio, fluidization temperature, bed depth, lime particle size, and the velocity in the bed. The variables of lime replacement rate and extent of calcium reaction between regenerations were tested in cyclic tests where the lime was cycled between fluidizing and regeneration conditions in the batch reactor.

These studies provided the basis for a number of guidelines and process correlations. The effects of bed depth (10 to 20 inches) and fluidisation velocity (4 to 8 ft/sec) were correlated as gas residence time in the bed, giving an approximately first order sulphur removal rate expression. Sulphur differential, the quantity of sulphur to which the lime was exposed in each gasification cycle, emerged as an important variable. As an approximation lime reactivity varied inversely with the square of this differential and increased directly with time to sulphur replacement ratio.

In parallel with the batch unit experiments, ERCA constructed a pilot plant in which the gasification and regeneration reactions could be studied under continuous operating conditions. A 10 million Btu/Hr water cooled boiler was included in the system to burn the gasifier product and dispose of the heat. Three tests designated CAFB Runs 1, 2, and 3 were made in this pilot plant during 1971. Run 1 lasted 230 hours of which 204 were at gasifying conditions. Denbighshire limestone (UK) was used throughout these runs together with Venezuelan fuel oil containing 2.5% sulphur. The pilot plant successfully demonstrated many features of the process including sulphur removal, lime regeneration, temperature control, start up, shut down, solids circulation, and release of the sulphur as a rich (8-10%) stream of SO_2. It also pinpointed areas for improvement which included reduction or elimination of carbonaceous deposits in cyclones and gas transfer ducts, minimisation of fines production and losses into the boiler, and improvement of regenerator oxidation selectivity.

CURRENT WORK OBJECTIVES

Work on this current contract constitutes the second phase of a six phase programme to demonstrate and evaluate the CAFB gasifier on a commercial scale with a power plant boiler. Work on this phase consists of three tasks.

Task 1 is operation of the CAFB pilot plant with the following set of objectives:

a. Verify that continuous gasification, sulphur and metal removal and lime regeneration results are as batch studies have indicated and evaluate the effects of bed depth, velocity, fuel/air ratio, lime make-up and fuel rate.

b. Determine minimum excess air requirements for operation of a continuous regenerator with good temperature control and maintenance of a low residual concentration of sulphur on the lime bed.

c. Demonstrate operability of the process over a prolonged period of time to show that accumulation of fines, agglomerates, carbon or other deposits do not interfere with continuous operation.

d. Demonstrate means of preventing or removing deleterious accumulations of tar or carbon from gasifier and transfer duct internals.

e. Determine effects of number and location of fuel injectors on gasification, sulphur removal, and carbon content of gasifier lime. Include operation with single oil injector passing through the air distributor.

f. Test and demonstrate means of process start-up, shut-down, turndown, and control. Determine maximum turndown ratio with independent control of gasifier and regenerator variables.

g. Determine effect of regeneration temperature on the maintenance of lime activity.

h. Study the existing burner operation with CAFB gasifier product. Establish operability with high gas velocity, measure flame characteristics, efficiency of combustion, production of NO_x, and flame stability.

i. Under conditions of lined out operation with equilibrated lime, measure SO_2 removal in the regenerator and determine rate of lime attrition and particle size distribution of solids carried over from the gasifier and regenerator. Determine engineering properties of equilibrium solids such as fluidized bed density, minimum fluidization velocity and particle size distribution.

Four pilot plant runs were planned to accomplish these objectives. To continue the numbering system begun in Phase I, these runs are designated runs 4,5,6 and 7.

Task II is an evaluation of additional limestones with two fuel oils in batch reactor experiments conducted between continuous unit runs. One of the test oils is to be high sulphur residue by-product of gas oil desulphurization. Originally, four new stones were to be studied. Because of factors uncovered during the first pilot plant run, the batch unit programme has been modified to include measurements of dust production tendencies of the stones under combustion conditions and to reduce the number of stones investigated to three.

Task III is a definition and assessment of the scope of engineering effort required to move the CAFB from the pilot plant stage through the development stage including the demonstration unit.

Tasks I and II are in progress at the Esso Research Centre, Abingdon, England. Task III was conducted by the Esso Research and Engineering Co., Florham Park, N.J., U.S.A.

REPORTING AND DISCUSSION OF RESULTS

During the performance of Task I the objectives for Task II were modified as a result of information generated during the first continuous gasifier run under Task I. Consequently, the reporting and discussion of results of Tasks I and II is set out in Section VI through IX in chronological order, so that the sequential logic of changes to objectives, equipment and techniques can be readily followed.

SECTION IV

EXPERIMENTAL EQUIPMENT

GENERAL

The experimental equipment used in this study consists of two batch reactor units and the continuous CAFB pilot plant. These units have been described (1) previously in detail. For the current work, the batch units remain essentially unchanged. However several modifications were made to the pilot plant on the basis of experience gained in the first three runs.

BATCH UNITS

Each batch unit contains a reactor, air and fuel systems, flare for product gas disposal, and gas sampling and analysis system as shown in the flow plan, Figure 1.

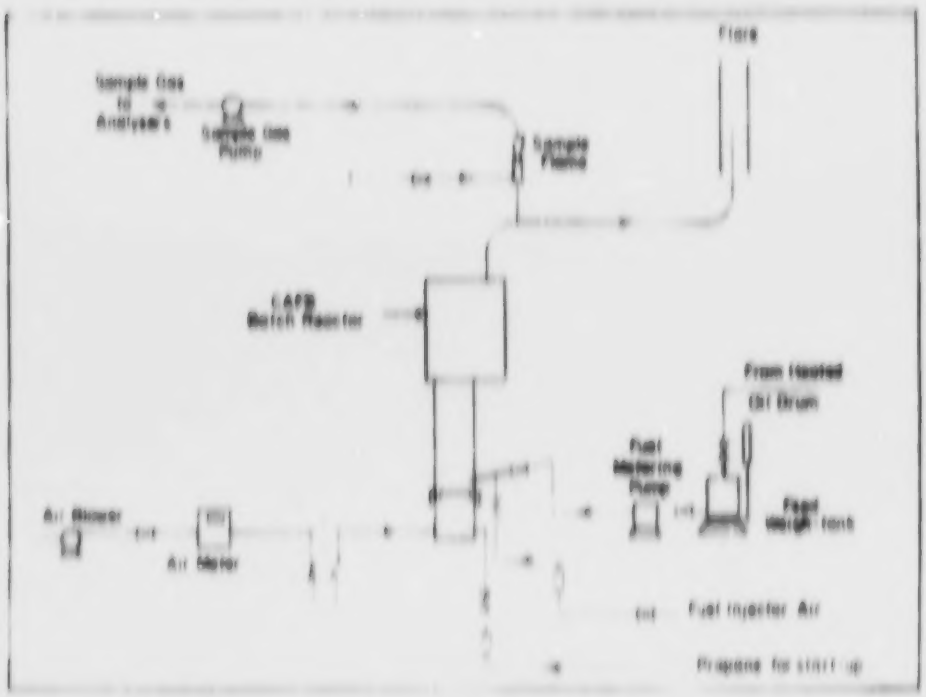

FIGURE 1 Batch Unit Flow Plan

A reactor is illustrated in Figure 2.

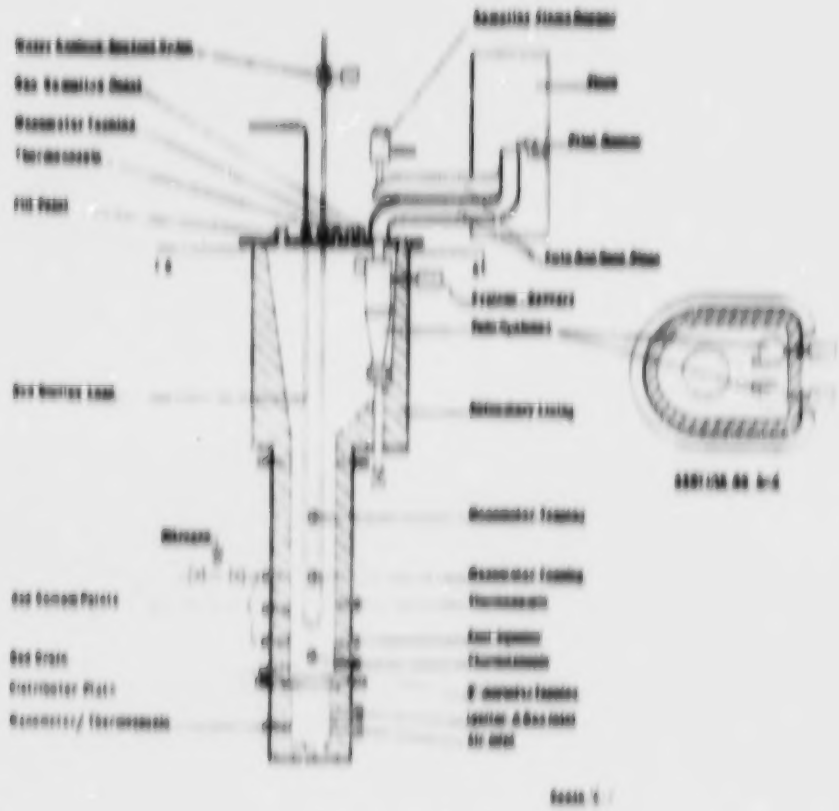

FIGURE 2 Batch Unit Reactor

The reactor is of refractory lined carbon steel construction. The lower section, which contains the fluid bed, is 7 inches I.D. by 33 inches high. The upper section is expanded to reduce gas velocity and is non symmetrical to permit internal cyclones to drain externally.

The plenum beneath the gas distributor is also refractory lined to serve as a combustion chamber for propane-air mixtures used during unit start up. The distributor is a "top hat" shape cast refractory design. The central raised section, five inches in diameter, contains 16 horizontal holes around its circumference.

For the current work, a rapper was installed on one cyclone in each reactor to prevent fine particles sticking to the cyclone walls. The pneumatic activator of the rapper is located outside the reactor and drives a striker rod through a gland to tap on the cyclone wall.

The gas analysis equipment used in this study is the same as used in Phase I, and is listed in Table 1:

Table 1

Batch Unit Gas
Analysis Equipment

Analyser	Type	Manufacturer	Model	Response	Range
SO_2	infra-red	Maihak	Unor 6	Continuous	0-1,000 ppm
SO_2	infra-red	Maihak	Unor 6	"	0-20% by vol.
SO_2	Conductimetric	Westoff	-	"	0-1000 ppm
CO_2	infra-red	Maihak	Unor 6	"	0-20% by vol.
CO	infra-red	Maihak	Unor 6	"	0-20% by vol.
O_2	Paramagnetic	Servomex	OA 137	"	0-25% by vol.

During fully combusting conditions and also regeneration of sulphids, the gas from the bed is monitored directly for CO_2, CO, O_2 and SO_2 and the appropriate information on combustion efficiency, sulphur removal efficiency or sulphur release deduced. During gasification, a portion of the product gas is burned in a sample flame located just above the reactor and the combustion products analysed for SO_2, O_2, CO and CO_2. The desulphurising efficiency of the gasifier is calculated from this analysis of the fully combusted gas. Because of the wide range of sulphur compounds present in the gasifier product itself, this is the only practical method for measuring gasifier sulphur removal efficiency.

CONTINUOUS PILOT PLANT

Process Flow Plan

Figure 1 is a process flow plan of the continuous pilot plant. The heart of the system is the gasifier-regenerator unit cast of refractory concrete contained in an internally insulated steel shell. The product gas of the gasifier fires a 10 million/ Btu/hr pressurized water boiler. The hot water is heat exchanged with a secondary water circuit which loses its heat through a forced convection cooling tower. The rest of the system consists of the necessary blowers, pumps and instruments to operate the gasifier, regenerator, burner and solids circulating system.

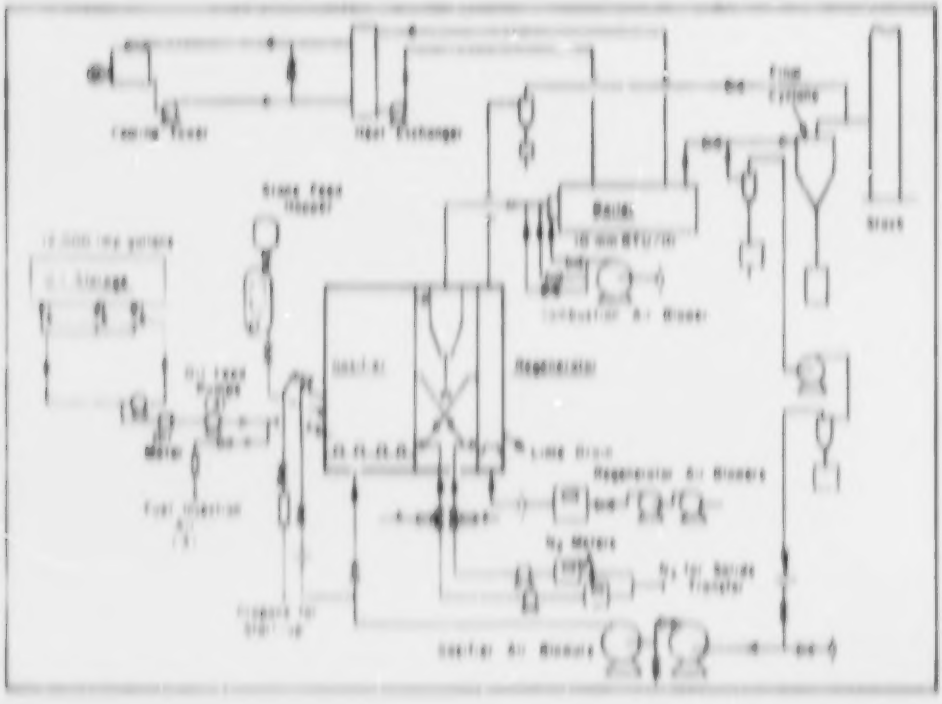

FIGURE 1 CAFB Pilot Plant Flow Plan

The gasifier itself sits within a pit to permit alignment of the gasifier outlet duct with the burner inlet. Fuel pumps, flow meters, and start up burner controls are mounted on a mechanical equipment console. Electrical instrumentation and manometers are mounted on a separate control console. Gasifier blowers are located in a separate blower house outside the main building, and the cooling tower is mounted on the roof.

The gasifier and regenerator reactors are cavities in a single refractory concrete block. The block contains other cavities which make up the gasifier outlet cyclones, the gas transfer ducts, and the transfer lines through which solids circulate between gasifier and regenerator. The gasifier cavity is rectangular in cross section, tapering from 17.5 x 37 inches at the distributor level to 19.5 x 39 inches at the 21 inch level; the upper portion has parallel sides. The regenerator tapers from 7" diameter at the bottom to 8" diameter 23" above distributor datum and remains parallel thereafter.

Modifications to Pilot Plant

A number of changes were made to the pilot plant between the end of Run 1 and the beginning of the current work phase. These modifications were intended to improve both working conditions and to overcome operating deficiencies encountered in the earlier runs.

Additional modifications made following Run 4 are discussed in a later section of this report dealing with preparations for Run 5. Those changes discussed here were performed in the period between December 1971 and July, 1972.

Control Cubicle :

The area including the control panel, analyser bank, and mechanical console, (pumps, flowmeters, etc) was walled off from the operating area containing the gasifier, boiler, and heat exchangers. This control cubicle was supplied with its own ventilation and heating system to provide a dust free area both to simplify maintenance of delicate equipment and to reduce dust exposure of personnel.

Dust Extractor :

A central dust extractor was installed in the main operating area with flexible ducts leading to areas of possible dust emission such as sample points, cyclone fines drains, and solids draw offs.

Limestone Feeder :

To avoid temperature upsets which had been experienced when limestone feed was interrupted to fill the original stone feed hopper, and to provide a constant measure of stone feed rate, a new stone feed system was designed and installed. In the new system limestone is fed to the gasifier through a gravity feed line from a pressurised weigh hopper. A vibrator is used to control the rate of stone addition. The feed line enters the gasifier through its side at a point below the bed level. A flow of air through the feed line keeps this line purged of gasifier product and helps move the stone into the bed. A ground level hopper is charged from bags of stone. A pneumatic transfer line moves the stone from the ground level hopper to an upper hopper from which stone is periodically dropped into the weigh hopper. The upper hopper acts as an air lock to avoid depressurising the weigh hopper.

Readout from the weigh hopper is continuously recorded on a chart from which both hopper content and rate of feed can be determined.

Regenerator Drain Control -

In the original control system, gasifier bed level was controlled by automatic adjustment of fresh limestone feed rate. Rate of solids draw-off from the regenerator was determined by manual adjustment of a vibrator on the regenerator drain line. Experience showed that regenerator solids drain flow was not dependable when throttled to a very low rate. Under low flow conditions the drain line would frequently block with solids. Also, when stone addition rate was controlled by gasifier bed level it was not possible to set stone feed to a predetermined rate. With the new limestone feeder it became possible to set stone feed rate at any desired level. A modification was therefore made to transfer control of gasifier bed level to the rate of regenerator solids draw off. The new regenerator drain system incorporated an air operated plug valve in the drain line, an adjustable cycle timer which controlled both the length of valve opening time and the frequency at which the valve opened, and a pressure switch operated by the depth of bed in the gasifier. When gasifier bed depth was below the set point, this pressure switch prevented the regenerator drain valve opening. When gasifier depth exceeded the set point, operation of the regenerator drain valve was under control of the cycle timer. With the cycle timer system, solids flowed from the regenerator in short, sharp bursts, which kept the drain line clear. Nitrogen injection into the drain line above the valve helped keep the solids free when the valve was closed.

Flue Gas Recycle -

A new blower was installed to boost pressure of the flue gas recycle stream returned to the gasifier air blowers. A more efficient cyclone was installed in the line to improve dust removal from the recycle flue gas.

Regenerator Pressure Control -

Good control of solids circulation in the gasifier-regenerator circuit requires that the pressure difference between the vessels be closely controlled. In the first three pilot plant runs this control was achieved by manual adjustment of a butterfly valve in the regenerator outlet line. This procedure was inconvenient and failed to provide completely satisfactory control. To improve matters, an automatic pneumatic instrument was installed to adjust valve position in response to the signal from a differential pressure measuring cell.

Regenerator Gas Outlet -

The regenerator gas outlet line was repositioned to enter the flue stack down stream of the external cyclone. This was done to remove the hazard of SO_2 release when solids were removed from the cyclone drain.

Air Distributors -

Both the gasifier and the regenerator air distributors were modified. The original regenerator air distributor was of the same design as used in the batch units, a cast refractory "top hat" shape with horizontal holes. After each of the first three pilot plant runs, a lump of agglomerated solids was found in the regenerator on the distributor. The shape of this lump indicated that it had originated in the defluidised region immediately above the centre of the "top hat" distributor. Also the refractory distributor cracked around its edge in each of the first three runs.

The refractory construction which led to the "top hat" shape is needed in the hot - units which employ propane firing beneath the distributor for start up. However conditions in the pilot plant regenerator are much less severe at this distributor, and stainless steel construction is possible. Therefore a stainless steel distributor was designed, tested and installed before Run 4. This distributor contains 5 stainless steel nozzles, each with 4 horizontal 1/8 inch i.d. holes. Hole spacing in the nozzles was selected to equalize air rate per unit area in the regions inside and outside of the nozzle circle.

Nozzles of the gasifier distributor were modified to reduce the attriting energy of the nozzle air jets. A shroud ring with offset, larger diameter holes, was placed around each air nozzle. Pressure drop through the inner holes of each nozzle still determines overall pressure drop and assures uniform air distribution. The inside surface of the shroud absorbs the jet energy however and permits a more gentle entry of the air into the bed.

Cyclone Outlet Tubes -

The central gas outlet dip tubes from the gasifier cyclones had been a trouble area during the first three runs. These tubes become coated with coke during gasification and are subjected to extreme conditions when the coke is burned off. The original gas outlets were simple stainless tubes which burned severely during Run 1. The tubes were omitted completely during Run 2 with consequent high dust loss rates.

Double wall stainless steel tubes used in Run 3 had provision for steam cooling during decoking operations. These tubes survived the run with slight damage. However, their thickness impaired cyclone operation.

Test specimens of various materials had been exposed to cyclone tube conditions during Runs 2 and 3, and self bonded silicon carbide was found to survive intact. New cyclone tubes of this material were therefore obtained and installed before Run 4.

Gasifier Outlet Duct -

In early runs, coke and lime deposits formed in the gasifier outlet ducts at points where the gas underwent sharp changes in direction. Major deposits formed at the junctions between the vertical and horizontal sections. After Run 1 the refractory lining of the bifurcated gas duct was recast to provide a smoothly rounded curve from the vertical to horizontal sections.

Gas Burner -

The gas flame reached nearly to the end of the primary fire tube of the boiler during the first set of runs. An attempt to shorten the flame by increasing the proportion of air added to the gas at the burner entrance was limited by increasing temperature within the burner throat. The burner was therefore redesigned after Run 3 to add a third air inlet point to increase mixing and turbulence at the gas outlet. Figure 4 shows the revised gas burner.

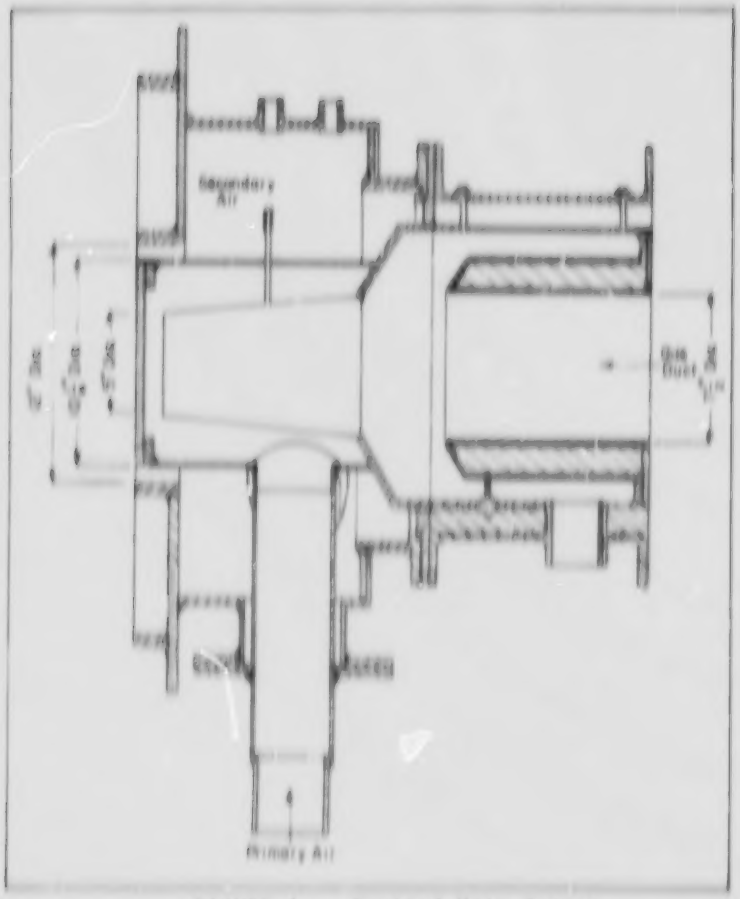

FIGURE 4 Revised Main Burner

Dampers and pitot tubes on the three air ducts leading to the burner permit individual adjustment of the rate to each point.

A new air blower was installed to replace the 1200 CFM blower supplied with the boiler. The new blower will permit a higher firing rate and can overcome the added back pressure of dust removal equipment added to the flue gas system prior to Run 3.

Boiler Probe

An air cooled probe was installed at the rear end of the primary fire tube to simulate a boiler superheater tube and test for the nature and extent of deposits or corrosion formed on such a tube. The probe is 1.5 inches i.d. by 39½ inches extension inside the boiler wall. Air for probe cooling is supplied by a diesel driven auxiliary compressor. A temperature controller which senses probe metal temperature by a thermocouple adjusts air rate to the probe. Tube temperature profile is recorded continuously by other thermocouples mounted along its length.

SECTION V

OIL AND LIMESTONE

Oil

The oils used in the present work have been heavy fuel oils from Venezuelan crudes obtained from Amuay refinery of Creole Petroleum Co. One supply, obtained in drums direct from Amuay has been used in batch unit tests. A second supply from bulk storage in the U.K. has been used in pilot plant work. An additional supply of very heavy residue has been obtained for batch unit studies which have not yet begun. This oil, also from Amuay, is vacuum pipe still bottoms produced when ordinary atmospheric pipe still bottoms is further distilled to give a vacuum gas oil which can be hydrofined to give a reduced sulphur fuel oil. Chemical and physical inspection results of these three oils are listed in Table 2.

Table 2

Properties of

CAFB Test Fuel Oils

		Amuay Batch Unit Tests	Amuay Pilot Plant Tests 1/12/73	14/3/73	25/7/73	Vacuum Bottoms
Property						
Specific Gravity		0.957	0.955	0.960	0.960	1.015
Kinematic Viscosity						
at 140 deg. F		201	221	-	254	-
210 deg. F		41.4	39.1	40.3	45.6	1180
280 deg. F		-	-	-	-	321
350 deg. F		-	-	-	-	71
Carbon	% by wt.	85.9	85.6	85.8	85.1	85.7
Hydrogen	%	11.1	11.4	11.6	11.1	10.1
Sulphur	%	2.35	2.43	2.42	2.41	2.95
Nitrogen	%	0.35	0.26	0.09	0.35	0.63
Conradson Carbon	%	11.6	10.9	11.1	10.8	17.4
Asphaltenes	%	7.1	4.8	5.3	6.0	8.9
Vanadium	ppm	166	343	370	315	530
Nickel	%	43	40	65	41	6
Sodium	%	38	35	37	38	9
Iron	%	3	4	-	4	108

LIMESTONE

Pilot plant runs have employed U.S. limestone BCR 1691 and a U.K. limestone from Denbighshire. Batch unit work has employed these two stones as well as U.S. stone BCR 1359. Two additional U.S. stones have been received. They are Tymochtee Dolomite and a limestone supplied by New England Electric Systems which we have designated NEES (1). This was selected because of its proximity to the proposed demonstration unit. No tests have been made on these two stones as yet. Chemical inspections of all the stones are listed in Table 1.

Table 1
Chemical Properties of Test Limestones
Limestone Composition

Stone:		BCR 1691	BCR 1359	Denbighshire	Tymoch-tee Dolomite	NEES (1)
Component						
CaO	% by wt.	45.6	54.1	55.2	31.1	55.0
MgO	"	3.35	0.60	0.30	20.9	0.5
SiO_2	"	13.65	0.75	0.68	3.1	0.8
Fe_2O_3	"	0.35	0.09	0.10	0.4	0.09
Al_2O_3	"	2.80	0.31	0.25	1.13	0.3
CO_2	"	35.7	44.0	43.4	43.6	43.2
S (Total)	"	0.44	0.12	< 0.05	0.13	0.03
Vanadium	ppm	41	50	26	< 25	< 20
Sodium	"	219	< 20	59	175	250
Nickel	"	40	30	21	10	50

SECTION VI

PILOT PLANT OPERATION - RUN 4

GENERAL

The experimental plan for Run 4 called for use of limestone BCR 1691 at a series of pilot plant conditions to test correlations based on batch unit studies. Limestone replacement rate, gasifier bed depth, and gasifier bed temperature were the major variables to be examined. A brief test of the effect of regenerator excess oxygen content was also included in the experimental plan.

A major goal of the study was to find if increasing gasifier bed depth would have the beneficial effect that batch studies had indicated to be possible. As events developed, properties of limestone BCR 1691 prevented accomplishment of these test goals and focussed attention on development of means to start up and operate with a stone which produces a great deal of dust under certain conditions.

The dust forming characteristics of this stone under full combustion conditions were found to be much worse than had been encountered with Derbyshire stone. This high rate of fines production caused a number of operating problems, extended the start up period, and eventually caused termination of the run after nine hours of gasification. Results of the post run inspection of the unit appear in Appendix B. Results and problem areas encountered in the run are described here.

GASIFICATION RESULTS AND DISCUSSION

Table 4 summarizes gasification conditions and results obtained. Initially a high lime replacement rate of 2.1 mole CaO was employed to build gasifier bed level. A slight reduction to 1.7 mole CaO was used during the final 4 hours. Sulphur removal efficiency of nearly 98% at the higher rate declined to about 91% when stone rate was reduced. However the gasification period was too short to consider these results to represent lined out conditions.

Table 5 lists the distribution of particle sizes in the solids from gasifier and regenerator beds and in solids recovered from the boiler fire tube and regenerator cyclone during gasification.

Table 5

Particle Size Analysis Run 4

Sample Location	Gasifier Bed	Regenerator Bed	Solids from Boiler Fire Tube	Regenerator Cyclone Fines (gasification)
Particle Size, wt % in Fraction				
1400 Micron +	39.8	21.6	19.1	.27 *
1400 - 1180	11.5	8.8	6.3	
1180 - 850	22.2	18.0	13.0	.27 *
850 - 600	15.3	13.9	13.0	
600 - 355	11.0	14.7	20.5	.27 *
355 - 250	0.5	5.8	7.1	.27 *
250 - 150	0.2	12.7	5.1	10.6
150 - 106	0.1	0.8	2.8	11.4
106 - Dust	0.5	3.7	12.0	76.9
Bulk Density g/cc	1.11	1.08	.85	1.03

* These particles had a white appearance; distinctly different from that of the finer particles.

The elutriation effect of the gasifier bed in removing particles smaller than the 355-600 micron fraction is apparent in Table 5. We would expect particles smaller than about 300 microns to be entrained at Run 4 test conditions. It is evident that little of the entrained material was returned to the gasifier by the cyclone. The presence of a wide spectrum of particle sizes in solids from the boiler fire tube also indicates poor cyclone performance. However the gasifier cyclone which drained back to the regenerator evidently was operating as there was an appreciable fraction of 150-250 micron solids in the regenerator bed.

The bulk density of the bed solids was higher than has been observed in earlier studies. Batch unit tests with ECR 1681 had given settled bed densities of about 0.85 g/cc compared with values over 1.0 observed here. A change in density of the fluidized bed has also been noted during the start up period of Run 4. This density increased from about 0.8 to nearly 1.1 during the start up. It is possible that a selective loss of lower density particles contributed to this increase in bed density.

The chemical analyses of bed samples listed in Table 6 show that silica content of the beds, and indeed all solids samples, increased over those of the raw limestones. This change indicates that minerals other than SiO$_2$ were preferentially lost from the system, probably as very small particles.

Table A

Desiccation Summary - Run 4

Day Hour	Temperature deg. C Desiccant	Regeneration	Superficial Air Velo ft/sec	Fluidized Bed Depth in.	Lithi Replacement Rel. Cat/Rel. S	Inc/Fuel % Smooth	Neighbor Removal %
1 0030	870	-	3.7	19.5	2.1	23.1	-
1 0130	882	1035	3.8	20.0	2.1	24.0	-
1 0230	872	1040	3.7	21.9	2.1	22.6	-
1 0330	870	1080	4.3	23.3	2.1	24.3	95.7
2 0030	875	1100	4.1	23.6	2.1	24.1	97.9
2 0130	881	1103	4.1	23.8	1.7	24.1	93.3
2 0230	875	1055	4.0	22.7	1.7	23.7	93.3
2 0330	872	1060	4.1	22.2	1.7	24.0	87.3
2 0430	872	1060	4.1	23.0	1.7	24.2	92.6

Table 4

Chemical Analysis of Lime Samples - Run 4

	Units	Gasifier (After Shutdown)	Regenerator (During Gasification)	Boiler (After Shutdown)	Regenerator (During Combustion)	Cyclone (During Gasification)
CaO	wt%	53.9	58.8	57.0	48.1	59.9
MgO	"	3.95	4.7	4.35	4.1	4.2
Al_2O_3	"	2.35	3.0	2.85	4.1	3.6
SiO_2	"	22.1	24.1	18.6	31.2	21.6
Fe	"	0.71	0.87	1.02	0.57	0.87
Na	"	0.06	0.07	0.07	0.06	0.08
K	"	0.13	0.15	0.11	0.08	0.08
Cl_2	"	1.84	0.73	0.36	0.21	0.12
S (total)	"	4.69	2.49	1.56	0.76	5.51
S as sulphate	"	4.39	2.06	1.28	0.71	2.58
Loss on Ignition	"	-	-	7.65	0.31	-
Gain on Ignition	"	0.21	0.16	-	-	4.09
SiO_2/CaO		0.41	0.41	0.33	0.31	0.36
MgO/CaO		0.074	0.080	0.074	0.080	0.070
Al_2O_3/CaO		0.043	0.051	0.050	0.080	0.060
SiO_2/CaO as Received Limestone			0.27			
MgO/CaO " " "			0.074			
Al_2O_3/CaO " " "			0.040			

The difference between gasifier and regenerator bed sulphur contents was 2.8% on stone indicating a good level of regeneration. A very high fraction, 99%, of the regenerator sulphur appeared as sulphate. This represents a considerably higher degree of sulphide oxidation than achieved in earlier runs and may indicate some oxidation of the sample during its collection. The regenerator cyclone fines show a slightly higher sulphur content than the gasifier bed sample. They also show a high content of sulphide which indicates that the fines passed through the regenerator without undergoing much reaction.

START UP PROBLEMS

Nature of Start Up Problems

During start up of the continuous pilot plant in Run 4, problems were encountered in the following areas:

(a) Blockage in solids transfer line
(c) Plugging in regenerator gas outlet system
(d) Dust emissions to boiler from gasifier
(d) Dust in flue gas recycle stream
(e) Dust emissions to atmosphere
(f) Regenerator Agglomerates

All of the problems were related to differences in the characteristics of stone mcb 1691 from those of the Pembrokeshire stone used in the continuous unit during Phase I studies [1]. The major differences are lower fusion temperature, the cause of problems (a) and (f) above and production of a higher proportion of very fine dust in a fluidised bed under fully combusting conditions, the cause of problems (b) through (e). The dust produced from mcb 1691 is more difficult to retain in collection equipment than that originating from Pembrokeshire stone. It also clings to surfaces of pipes, cyclones, control valves etc, and is difficult to dislodge without application of direct mechanical force. It does not drain from hoppers, or even vertical pipes, without continuous rapping.

Transfer Line Blockage

Heatup of the pilot plant for Run 4 started on July 28. Stone addition was begun the afternoon of August 1, and by early on Aug 2, a hot fluidised lime bed was established under kerosene combustion conditions. However, efforts to establish good solids circulation through gasifier and regenerator were unsuccessful. Attempts to seal out the transfer lines did not improve circulation very much. The unit therefore was shutdown on August 4, allowed to cool and opened for inspection on August 8. The blockage was found to consist of a fused mass of lime particles which obstructed most of the mixing pocket in the regenerator to gasifier (R to G) transfer line. The transfer line from this mixing pocket to the gasifier also contained a quantity of material with the appearance of foamed slag.

Reconstruction of the startup procedure indicated that the obstruction was caused by limestone particles and fines entering the R to G transfer line during the initial stages of stone addition while the pocket and transfer line were heated by direct gas flame. The geometry of the system is now such that stone enters the bed from the stone feeder at a point directly opposite the R to G line. The start up burner is between the stone feed point and the R to G line. During start up, gasifier pressure had been maintained well above regenerator pressure to drive hot gas into the regenerator to raise its temperature. It is evident now that flame had actually entered the transfer line along with stone. The silica content of the WCR 1691 stone lowers its melting point enough to allow fusion under these conditions, producing, in effect, flame spraying of fused stone directly into the R to G transfer line.

The solution to this problem is to adjust pressure balance during the early stages of stone addition to avoid overheating the R to G transfer system. This method was adopted for the second start up. As added precautions, a thermocouple was installed in the transfer pocket, and low silica Denbighshire stone was added initially to form a bed deep enough to cover the transfer line. The experience gained during the second bed addition on 15 August indicates that flow of hot gas and stone into the R to G line can be prevented by adjusting the pressure balance and that no trouble would have been experienced with the 1691 stone alone. The thermocouple in the transfer pocket is a valuable guide to temperature and flow conditions at that point, and its continued use is recommended. No blockage was encountered during the second stone addition, and once good fluidisation and combustion were established, high circulation rates between beds were easily obtained. Some difficulty was encountered in establishing initial fluidisation with 100 to 1200 micron stone. Indeed each startup had encountered some trouble during the initial period of stone addition because of the high heat load required for stone calcination and the high gas velocity required to fluidise uncalcined stone. In this startup, good fluidisation and operating bed temperature were achieved after addition of some precalcined stone removed from the unit in Run 3. Use of calcined lime is recommended for startup in future runs.

Regenerator Gas Outlet System

During the startup per' ' of run 4, dust in the regenerator off gas stream continually b, ked the regenerator cyclones and the gas exit line and control valve on stream of the cyclone. It was possible to keep the cyclone functioning only by continuously rapping on its body. Without this rapping, solids failed to drain, quickly filled the cyclone interior, and went overhead to form restrictions further down stream. It was found that the overhead line remained relatively clean in straight sections and smooth bends but plugged at sharp bends and fittings.

The character of regenerator fines changed when gasification began on August 20. Almost immediately the regenerator cyclone became free draining and operative without rapping.

The colour of the fines also changed to a darker hue. Microscopic examination showed the dust from the combustion period to have a high proportion of very fine particles.

Fig. 5A Photomicrograph of Regenerator Fines Under Combusting Conditions.

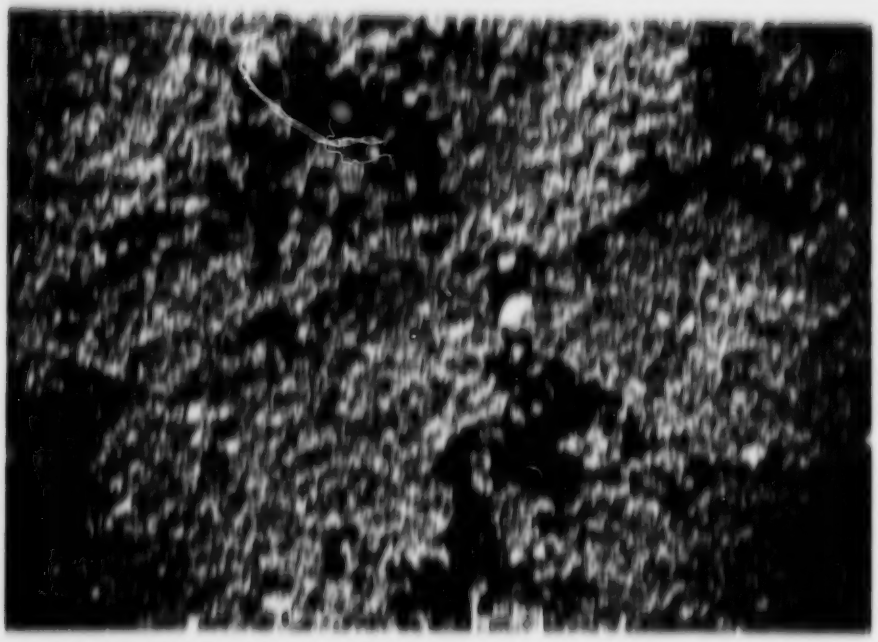

Fig. 5B Photomicrograph of Regenerator Fines Under Gasifying Conditions.

Figure 5 shows photomicrographs of regenerator solids obtained under combusting conditions (5A) and gasifying conditions (5B). The larger particle size and reduced agglomeration tendency of the gasifying samples is apparent.

Nevertheless, the regenerator outlet control valve eventually plugged after nine hours of gasification and the run was terminated. Inspection of the outlet system revealed no accumulation of fines except at the valve itself. The solids forming the plug appeared to be more like those formed during combustion than during gasification, and it is possible that they were a remnant of the pregasification period which had dislodged from the transfer line and moved down stream to the valve.

Dust Emissions to Boiler

A large quantity of dust passed from the gasifier into the boiler. Much was retained within the boiler, particularly at the back end of the main fire tube, where the flue gases change direction abruptly through 180 degrees, but some passed through to be caught in the external cyclone or to escape from the stack. There are indications that fines losses decreased during gasification, but the period was too short to confirm this observation.

Unlike previous operations where fines also entered the boiler, this time we were unsuccessful in withdrawing solids from the drain points at the boiler end. The solids failed to drain because of their steep angle of repose.

A contributing factor to high losses from the gasifier was the greater bed pressure drop used in the current run. In earlier runs, a maximum bed pressure drop of 19 inches water gauge (w.g.) was used. In the current run this was increased to 26 in w.g.

This increase in pressure drop together with a reduced density of fluidised cyclone fines (due to lower average particle size) caused level of solids in the cyclone drain line to reach the cyclone itself and decrease cyclone efficiency.

Dust in Flue Gas Recycle

Two stages of cyclones in the flue gas recycle system failed to remove fines to a degree which would assure long term cleanliness of the gasifier distributor. No pressure drop increase in the distributor was observed during the short test period, but fines observed in the flue gas sample line filter downstream of the cyclones indicated that a problem eventually would have occurred.

External Dust Emissions

The external settling chamber and cyclone which proved adequate for final flue gas clean up in Run 7 was unable to cope with the fine dust produced under combustion conditions in the current run.

In part, poor performance of the external cyclone was also caused by the sticky nature of the fines. The interior of the cyclone was quickly coated with a layer of solids which impaired its efficiency.

Regenerator Agglomerates

Analysis of Run 4 temperature records and inspection of samples retrieved from the regenerator revealed an additional problem with BCR 1691 stone. During the early portion of the Run 4 gasification period, an upset in pressure balance interrupted solids circulation and allowed a brief temperature excursion. The temperature in the lower portion of the bed reached 1110 deg. C. After this upset, the temperatures in the upper and lower regions diverged with the lower temperature lagging about 80 deg. C. below the upper one. This condition indicates poor fluidization. When the regenerator was opened for inspection a number of agglomerates were found. We believe that these lumps formed during the brief high temperature period in spite of bed fluidization. In previous runs with Denbighshire stone there had been temperatures of over 1110 deg. C. without encountering similar losses in fluidization. This difference in behaviour is attributed to the lower fusion point of the lower purity BCR 1691 stone.

SECTION VII

BATCH UNIT STUDIES

GENERAL

In the original programme of work, batch unit studies were to determine the suitability of additional fuel-limestone combinations for CAFB applications. However since Run 4 operations revealed a serious problem with BCR 1691 stone which had not appeared in earlier batch unit tests, the batch programme was revised to include an investigation of dust forming tendencies under a variety of operating conditions.

In earlier batch work there had been a comparison of dust losses between stones during gasification-regeneration cycles. No such comparison had been made under fully combusting conditions. In the normal batch unit test procedure there had been little exposure of the solids to combustion conditions in the absence of sulphur except during calcination. Consequently the conditions employed during Run 4 start up produced an entirely unexpected result in that the BCR 1691 stone formed copious quantities of a dust with a very sticky nature.

The batch unit test programme was revised to accomodate the following objectives:

- Determine if continuous unit conditions which produced large quantities of sticky dust could be duplicated in batch units.
- Compare dust producing tendencies of Denbighshire and BCR 1691 stones under different conditions.
- Provide a quantitative measurement of dust production to be expected under start up and operating conditions with Denbighshire, BCR 1691, BCR 1359, and two additional stones to be provided by New England Electric System (NEES).
- Measure sulphur absorption performance of BCR 1359 and the two NEES stones.
- Conduct tests of the feasibility of operating CFB with vacuum pipe still bottoms.

BATCH UNIT PROCEDURES

Several test procedures were employed to measure sulphur absorption and dust producing characteristics of the stones. These included fresh bed tests in which a new batch of calcined lime was used for each test, cyclic gasification tests in which a single batch of lime was cycled between gasification and regeneration conditions, and kerosene and fuel oil combustion tests in which the appropriate fuel was burned in the fluid bed with excess air to control temperature.

Start Up and Calcination

The bed is calcined by combustion of propane below the distributor and by direct injection of kerosene into the bed. First of all, the bed space temperature is raised to 850 deg. C. by gas combustion below the distributor. Then, 4000g of limestone is added. This is heated to 750 deg. C. using gas before switching to direct kerosene injection. Due to the strongly endothermic nature of the calcination reaction, the temperature remains in the region of 800 deg. C. until CO_2 evolution ceases. When the temperature rises to 850 deg. C, indicating that calcination of the original charge is complete, 250G. limestone is added and calcined. Further batches of this size are added and the procedure repeated until the target bed depth is reached.

Combustion Test

For a test with kerosene combustion, the injection of kerosene is continued with the bed past the point of complete calcination. For fuel oil combustion, the fuel supply is simply switched to heavy fuel oil from its heated supply drum. The fuel rate is set to give a bed temperature slightly in excess of the target, and fine adjustment is carried out with the aid of a cooling coil. Relevant data on gas analysis and unit behaviour are recorded and appropriate bed and cyclone samples taken. Combustion is continued for the required length of time.

Gasification Test

To achieve gasification, the fuel supply is changed from kerosene to fuel oil when calcination is complete, and oil rate is increased to obtain an air/fuel ratio of about 25% of stoichiometric. The sample flame burner and external flare are lit and the gas analysers connected. Relevant data are recorded and bed samples and cyclone samples taken at prescribed intervals. Gasification is continued for the requisite length of time.

Regeneration

Regeneration of the sulphided stone from a gasification cycle is performed by stopping the fuel supply and continuing the flow of air. Oxidation of carbon and CaS in the bed raises temperature to the regeneration level. Relevant temperature and analytical data are collected, and a bed sample is taken when regeneration is complete.

Shut Down

At the end of a run, the bed temperature is allowed to drop to 700 deg. C before the bed is removed through the drain point just above the distributor. Draining the unit is much easier when the bed is hot since the solids flow better under these conditions. When all the bed has been removed, all ancillary equipment is shut off.

Cycle tests

Cyclic tests are the nearest simulation to continuous gasifier operation that can be obtained in batch units. The same charge of lime is subjected to repeated cycles of sulphur absorption and regeneration. After each regeneration a portion of lime is removed and replaced by an equivalent amount of fresh limestone. The limestone calcines to lime during the early part of the next gasification cycle. Without replacement, the activity of the lime bed gradually declines. With replacement, the activity falls initially, but in a few cycles lines out at an equilibrium level which is influenced by the rate of replacement.

Enough cycles are performed at each set of operating conditions to establish the lined-out sulphur removal efficiency for those conditions.

Cyclic tests use the same calcination and start-up procedures as the fresh bed tests. However, after the initial start, a series of gasification and regeneration cycles follow each other. Sampling and gas analysis procedures also are the same as used in fresh bed tests.

When a regeneration cycle is complete, the fluidisation cools the bed very rapidly. Since solid sample withdrawal is very difficult in the absence of fluidisation, it is not possible to draw a regenerated bed lime sample before temperature drops below the level needed for the next gasification cycle. Therefore the following procedure was adopted.

(1) When regeneration was complete as indicated by end of SO_2 emission and fall of bed temperature, fluidising air was stopped.

(2) Replacement limestone was added.

(3) Fluidising air was resumed and when temperature reached the desired point, oil feed was resumed.

For the purpose of stone comparison, the target conditions in each test were the same. These conditions are listed below:-

Air/fuel Ratio	(% of stoichiometric)	20
Gasification Temperature	(deg. C.)	870
Make-up rate	(wt CaO/wt S)	1.6
Bed Depth	(inches w.g.)	15
Gas Velocity	(ft/sec)	6
Potential Sulphur Differential	(% wt)	3
Limestone Particle Size	(microns)	600-1175

BATCH UNIT PROGRAMME

The batch unit test programme for Phase II is summarised in Table 7. The design of this programme with regard to fines production rates and properties was based very much on our experience in continuous pilot plant run 4. There, fines produced from stone BCR 1691 under fully combusting conditions did not drain freely from cyclones, whereas those produced under gasifying conditions did.

Table 1

Test Programme for 7075 Series Alloy

Test	Cycle	Test	Conditions	Objective
1 — A	RCR 1255	Corrosion	Calculation and prolonged immersion at 45°C max. C	Measure fibre loss rate and proportion of test produced with RCR 1255 cycle
1 — B	—	—	Take at 1255 max. C	—
1 — C	—	Wheel 1.2m x Test rig	Calculation and immersion at 45°C max. C	Same
1 — A	—	—	Calculation and localization at 45°C max. C	Same
1 — D	—	—	Localization — temperature cycles	Measure fibre and sulphur material efficiency and fibre test rig
1 — F	—	Corrosion	Thirteen immersion cycle cycle and fibre Test rig	Measure fibre loss rate during and some immersion on traditional fibre test
1 — C	—	Wheel 1.2m x Through Tensile	Localization — temperature fibre	Measure fibre and sulphur material efficiency and fibre loss test with real tensile load
1 — A through 1 — F	Combination	Corrosion and Wheel 1.2m x 0.1m	Option g, varies take on in Test 1 service only F	Measure fibre loss rate under different sets of conditions with comprehensive view
2 — A	RCR 1255	Corrosion	Calcium fibre and prolonged immersion at 45°C max. C	Measure fibre loss rate and fibre proportion with RCR 1255
2 — B	—	Wheel 1.2m x Test rig	Calculation and prolonged localization at 45°C max. C	Same
2 — C	—	—	Localization — temperature cycles	Measure fibre and sulphur material efficiency and fibre loss rate
4 — A	RCR 1255	Corrosion	Calculation and prolonged immersion at 45°C max. C	Measure fibre loss rate and proportion of test produced with RCR 1255 cycle C
4 — B	—	Wheel 1.2m x Test rig	Same	Same
4 — C	—	—	Localization — temperature cycles	Measure fibre and sulphur material efficiency and fibre loss rate
5 — A	RCR 1255 C	Corrosion	Calculation and prolonged immersion at 45°C max. C	Measure fibre loss rate and proportion of test produced with RCR 1255 cycle C
5 — B	—	Wheel 1.2m x Test rig	Same	Same
5 — C	—	—	Localization — temperature cycles	Measure fibre and sulphur material efficiency and fibre loss rate.

It was suspected that the higher resistance to flow of the combustion fines was due to their containing a much higher proportion of very fine particles. Why less of the very fine particles should be produced during continuous gasification and regeneration was not clear. It was considered possible that the presence of sulphur on the stone, the higher regenerator bed temperature, an ageing effect or a combination of these could be the answer. Any of these changes could have altered the particle surfaces in such a way as to make them less susceptible to decrepitation.

Work on this programme has now been completed to test IC except for test 10 with Amuay vacuum bottoms.

EXPERIMENTAL RESULTS

The results to date permit a comparison of sulphur removal efficiencies, fines production rates, and fines properties from three limestones, BCR 1691, Denbighshire, and BCR 1159.

Sulphur Removal Efficiency

Sulphur removal efficiencies (SRE) for the three stones were compared by cycle tests. Adverse operating conditions were chosen so that SRE would be less than 100%. When successive cycles of gasification and regeneration are repeated, SRE, under adverse operating conditions, falls for several cycles and then lines out. Comparison values of SRE are measured at the lined out level. Detailed results of the tests are listed in Appendix A. Results are summarised in Table 8.

SRE's are similar for the three stones. However, although the target conditions were the same for each test, actual conditions did vary slightly as shown in the table. Consequently, for comparison, lined out SRE was calculated for each of the test conditions for each stone from the equation derived for BCR 1691 [1]. From the ratio of measured to calculated SRE it appears that the Denbighshire stone is the most active and BCR 1691 the least active. BCR 1159 is intermediate in activity but appears to be closer to 1691 than Denbighshire.

Fines Production Rate

The measurement of fines production rate was based on bed losses. Table 9 summarises the bed loss results for the three limestones under a variety of CAFB conditions. Detailed results are given in Appendix A.

With BCR 1691, highest bed loss rate occurred during kerosene combustion with a fresh bed at 870 deg. C. Loss rates during fuel oil combustion and gasification where sulphur was being absorbed by the stone were lower, those during gasification where sulphur absorption was more rapid being less than during combustion.

Table 3

Summary of Bench Unit SRE Results

Luminaire	Persistence Time (sec)	P.S.D.+ (at ½)	Take-up Cart/S wt.	Take-up Rate Rule	SRE (Measured) %	SRE * (Calculated) %	Ratio of SRE (Meas'd.) to SRE (Cal.)
BCR 1641	0.20	1.96	1.59	.41	74.5	78.2	0.95
BCR 1259	0.15	1.67	1.61	.52	75.0	76.0	1.0
Retrolighting	0.17	2.18	1.49	.65	76.0	64.9	1.17

* Calculated from equation for predicting SRE's for BCR 1641.
\+ Projected Halphen Differential

Table 4

[illegible]

Test No.	Test Condition	Lubricant	Temperature, °F (°C)	Total Cycles to Failure	Cycles Loss from Test, Groups 1 Through 3	Cycles Loss from Test, Groups 4 Through 6	Average Loss Rate over 6 Tests, %/1000
1-A	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
1-B	[illegible]	"	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
1-C	[illegible]	"	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
1-D	[illegible]	"	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
1-E	[illegible]	"	[illegible]	—	—	—	[illegible] *
1-F	[illegible]	"	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
2-A	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
2-B	[illegible]	"	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
2-C	[illegible]	"	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
2-D	[illegible]	"	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
2-E	[illegible]	"	[illegible]	—	—	—	[illegible] *
2-F	[illegible]	"	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
3-A	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
3-B	[illegible]	"	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
3-C	[illegible]	"	[illegible]	—	—	—	[illegible] *

* Figure value is loss rate during 1st cycle, overall vs Loss not loss rate after 1 cycles.

the sulphided and aged stone from the cycle tests also showed a relatively low loss rate during kerosene combustion at 870 deg. C. The lowest loss rate measured by tests of the fresh bed type, however, was obtained during kerosene combustion at 1050 deg. C, the temperature level used in regeneration. This was only slightly above the lined out rate for the cycle tests in which the rate fell from 19.8 g/min in the first cycle to a stable level of 4.5 g/min after 5 cycles. We can conclude, therefore that raising bed temperature or introducing sulphur into the bed decreases BCR 1691 loss rate. This explains the reduction in losses observed when gasification was commenced in Run 4. Sulphur had been absorbed by the bed which was also being subjected to the high temperatures of regeneration. Prior to this, no sulphur had been introduced to the bed since kerosene combustion had been employed, and the regenerator temperature was low since no reaction was taking place.

A bed ageing effect is also indicated. Hourly losses during fresh bed tests decreased as the tests proceeded whilst loss rate during cycle tests had fallen from 19.8 g/min to 4.5 g/min by 5 cycles. It is important to note, however, that ageing during fresh bed tests is distorted to some extent due to the fact that bed depth was decreasing during each test. That bed depth has an important effect on loss rate was established in the Phase I work [1].

The relative importance of all variables effecting bed loss rate of BCR 1691 is summarised in the equation below. This was derived from further analysis of the fresh bed test results.

$$L = \frac{221.1 \; S \; D^{2.17}}{A^{0.44} \; S \; (T-750)^{1.80} \; S^{0.41}}$$

L = Loss Rate (g/min)
D = Bed Depth (inches)
A = Bed Age (hours)
T = Bed Temperature (deg. C)(750 deg. C is taken as $CaCO_3$ decomposition temperature).
S = Bed Sulphur Content including inherent sulphur (% by weight)

This equation shows the approximately square relationship between losses and bed depth as observed previously. The loss rate of 1.1 g/min calculated from the above equation for cycle test conditions is in fair agreement with the measured rate of 4.5 g/min.

With Denbighshire stone fresh bed tests showed a decrease in loss rate with increased temperature and an increase when fuel was used instead of kerosene. The stable rate during cycle tests was lower than for fresh bed tests. The lowest rate, however was measured with aged,sulphided stone under kerosene combustion.

In light of our experience with the two previous stones, only kerosene combustion, fuel oil gasification and cycle tests were studied with BCR 1159. The gasification fresh bed test gave a higher loss rate and cycle tests a lower rate than the fresh bed test with sulphur free kerosene combustion.

Table II

Summary of Bench Unit Loss Tests

Loss Rate (g/hour) *

Conditions:	Kerosene** Combustion 870 deg.C	Kerosene** Combustion 1070 deg.C	Fuel Oil** Combustion 870 deg.C	Fuel Oil** Gasification 870 deg.C	Kerosene Combustion in Sulphided Aged Bed 870 deg. C	Gasification Regeneration Cycles
Limestone						
BCR 1691	22.6	5.7	13.5	7.8	8.5	4.5
BCR 1759	2.1	-	-	3.6	-	1.9
Dealumination	4.1	1.1	4.1	4.7	1.8	1.4

* At all conditions except gasification/regeneration cycles, loss rate has been calculated over the first four hours of the test. The cycle loss rate is the stable loss rate.

** Fresh Bed Tests

The bed eyeing effect which decreased losses with BCR 1691 was also evident with Denbighshire and BCR 1159. For example, within 5 cycles, loss rate dropped from 6.6g/min to 1.4g/min with Denbighshire and from 6.4 g/min to 1.9 g/min with BCR 1159.

These results reveal several interesting comparisons between the stones. These are summarised in Table 10.

Neither of the higher purity stones gave the very high loss rate found with BCR 1691 during fluid bed combustion with kerosene at 870 deg. C. Both Denbighshire stone and BCR 1691 gave lower loss rates during kerosene combustion when temperature was increased from 870 deg. C to 1050 deg. C.

Although the temperature effect was less dramatic for Denbighshire, its loss rate remained below that of BCR 1691. No high temperature test was made with BCR 1159.

All three stones exhibited a decreasing loss rate with age during the initial test periods. The change became less significant at long exposure times. This age effect may be due to a strengthening of particles by a sintering process, to elimination of particles of lower initial strength, or to a combination of these factors. The significant decrease in loss rate at the higher temperature appears to be a consequence of the more severe sintering which would be expected under those conditions.

The effect of changing from kerosene to fuel oil differed between the low purity and high purity stones. Whereas with BCR 1691, fresh bed loss rate decreased when fuel oil replaced kerosene combustion and decreased again on going to fuel oil gasification, opposite directional results were found with Denbighshire stone. Results from the shorter test programme on BCR 1159 indicate that its behaviour is similar to that of Denbighshire. In spite of its high loss rate during fresh bed gasification, the Denbighshire stone gave the lowest rate of the three during gasification - regeneration cycles.

We believe that loss rate differences between kerosene and fuel oil operation are due to sulphur in the oil. However, the mechanism of the sulphur effect must be complex to increase losses with the pure stones whilst decreasing losses with the lower purity BCR 1691.

In addition to the results already discussed, bed losses were also measured during calcination and are summarised in Table 11.

Table 11

Summary of Batch Unit Calcination Losses

Stone	Low Sulphur Fuel Kerosene and propane *	High Sulphur Fuel 2.34 S Oil *
BCR 1691	18	15
Denbighshire	16	24
BCR 1359	6	-

* Losses as % of calcined stone

The low calcination loss rate from stone BCR 1359 makes it particularly attractive. The effects of sulphur on loss rate in calcination of Denbighshire and BCR 1691 stones are in the same direction as observed in the fresh bed tests with these stones.

Nature of Fines

Information on fines properties was gathered from material collected in the cyclone and deposited in pipes downstream of the cyclone.

Throughout the tests the cyclone operated satisfactorily with the aid of a mechanical rapper, and regular samples of fines were obtained. Table 12 contains the results of a microscopic examination of these fines together with cyclone efficiencies. With respect to physical appearance, the fines have been separated into five groups. Examples from each group are shown in Figures 6 to 10. The only fines represented by Figure 6 are those collected during kerosene combustion at 870 deg. C in BCR 1691. Most of the particles appear to be less than 50 μ. The fine particles seem to be adhering to each other to form loose agglomerates and to the surface of the few larger particles that are around. This stickiness was also observed during kerosene combustion in Run 4. Figure 7 shows the type of particles obtained when BCR 1691 was subjected to kerosene combustion at 1050 deg. C, to fuel oil combustion at 870 deg. C, and to gasification-regeneration cycles followed by kerosene combustion at 870 deg. C. There appears to be some larger particles in this group and there is less evidence of the stickiness. In Figure 8, the type of fines collected in all instances where gasification or gasification/regeneration cycles were carried out are illustrated. Due to the presence of carbon, it is difficult to estimate the particle size range present. However, it would appear that the great majority of particles are less than 50 μ. The fines collected under combustion conditions with Denbighshire and BCR 1359 stones are represented in Figure 9.

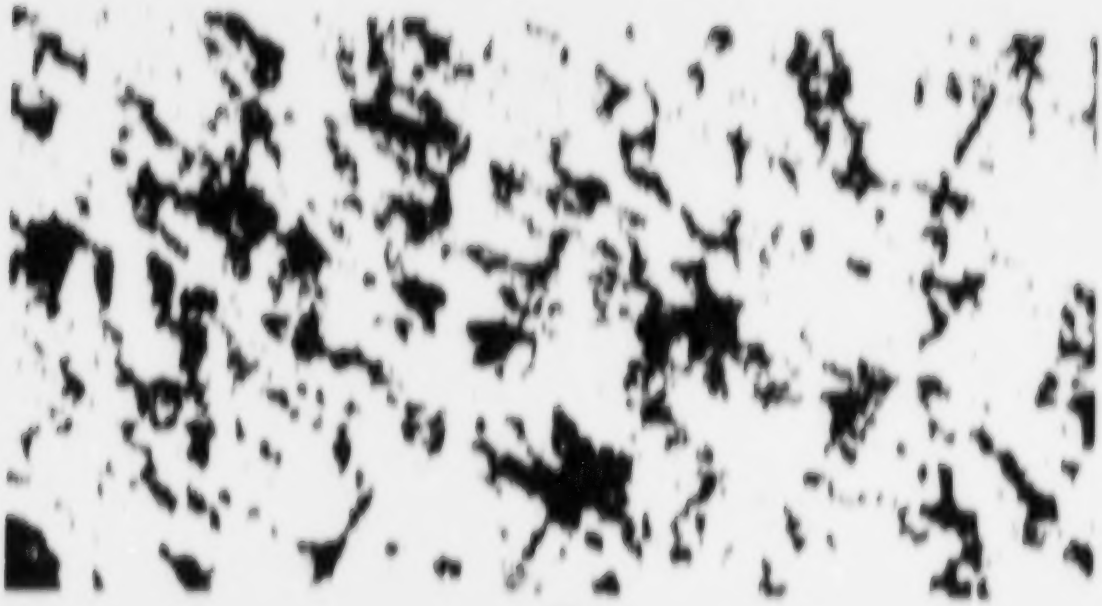

Figure 6 Cyclone Fines - BCR 161I (Kerosene Combustion 870 deg. C.)

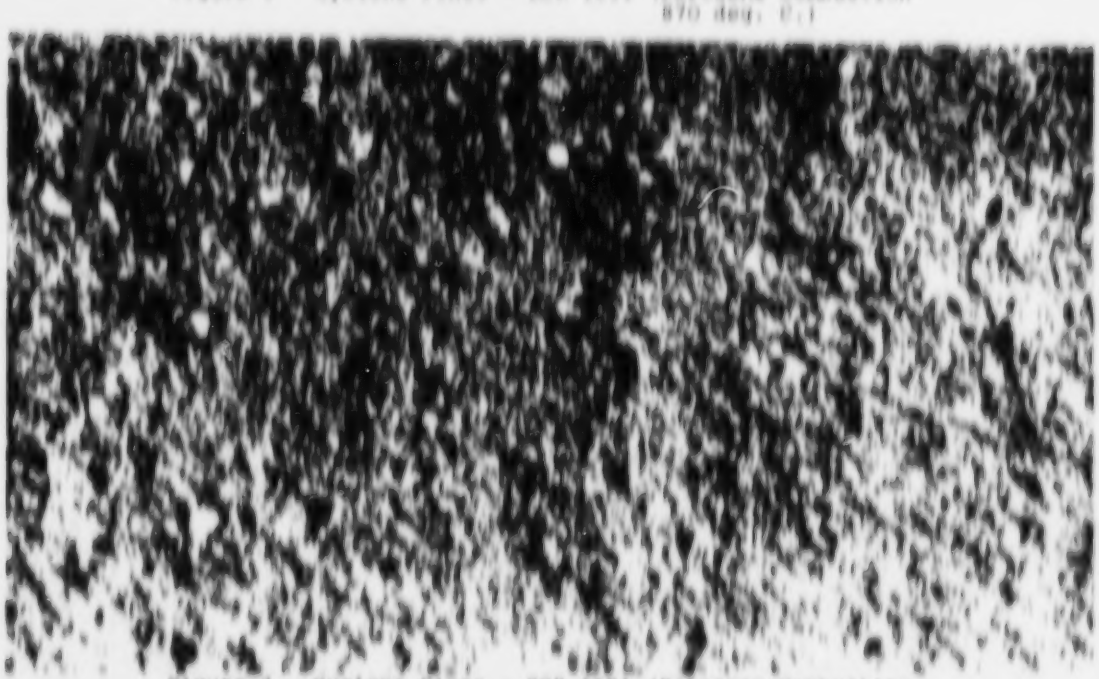

Figure 7 Cyclone Fines - BCR 161I (Kerosene Combustion 1050 deg. C, Fuel Oil Combustion 870 deg. C., Kerosene Combustion 870 deg. C on aged bed).

Figure 8 Cyclone Fines - BCR 1691, BCR 1359 and Denbighshire Gasification - Regeneration Cycle

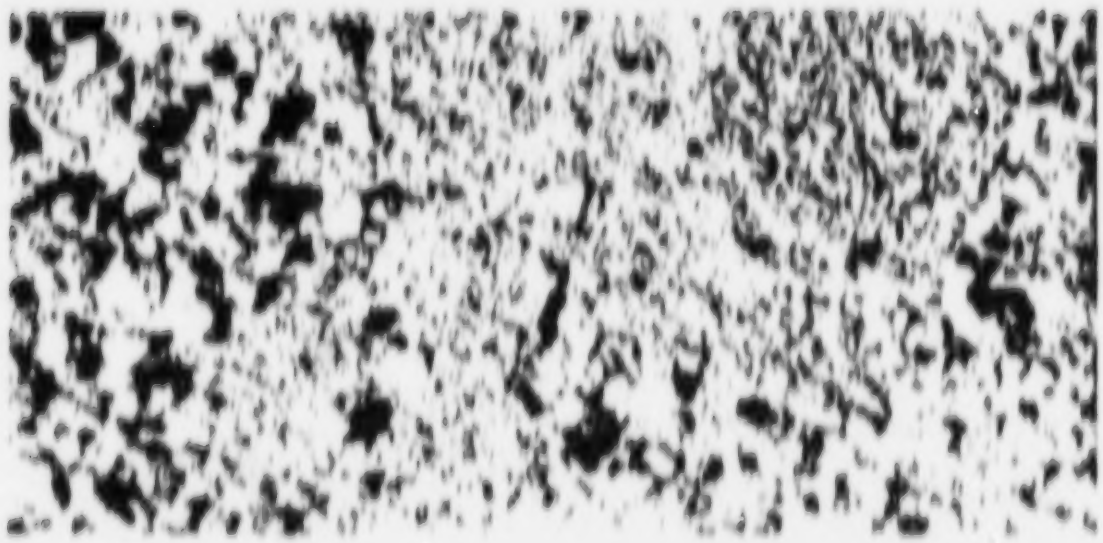

Figure 9 Cyclone Fines - BCR 1359 and Denbighshire (All combustion conditions).

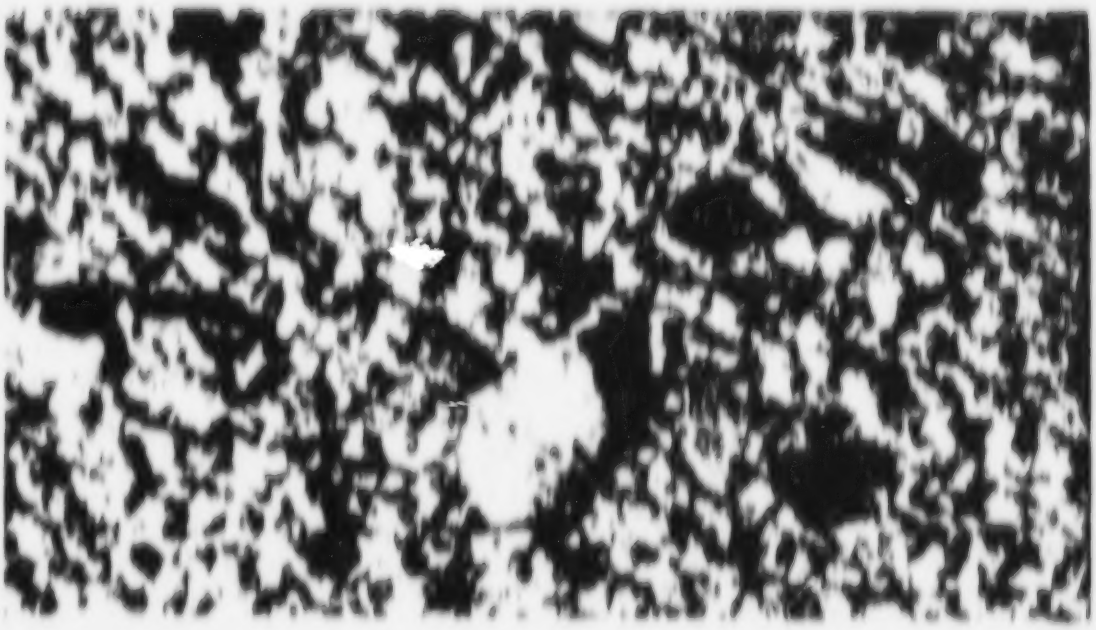

Table 12

Values of Cyclone Flows and Bench Test Problem

Test Condition	Cyclone	Comments	Cost	Cyclone Efficiency, %
Rect. cyclone at 470 deg.C	610-1400	Lower cyclone flow is of particles less than 50 u	min. 50 u	80
Rect. cyclone at 1250 deg.C	"	Mainly discrete particles less than 50 u	min. 50 u	100
Fluid bed cyclone at 470 deg.C	"	Mainly discrete particles less than 50 u	min. 50 u	72.4
Fluid bed gasification 470 deg. C	"	Mixture of carbon & lime	-	68.1
Calcination/Regeneration Cyclone	"	Mixture of carbon & lime	-	62.3
Rect. cyclone-multiclone used bed 470 deg. C	"	Mainly discrete particles less than 50 u	min. 50 u	94.1
Rect. cyclone at 470 deg.C	Calculations	Discrete fraction of particles up to 1000 u	>1000 u	76.4
Rect. cyclone at 1250 deg.C	"	"	>1000 u	62.3
Fluid bed cyclone 470 deg. C	"	"	>1000 u	68.2
Fluid bed gasification 470 deg. C	"	Mixture of carbon & lime	-	76.4
Calcination/Regeneration Cyclone	"	Mixture of carbon & lime	-	100
Rect. cyclone - multiclone used bed 470 deg. C	"	Mainly particles smaller than 50 u	min >50 u	4 5.4
Rect. cyclone at 470 deg. C	Lab Tests	Discrete fraction of particles up to 1000 u	>1000 u	85.1
Fluid bed gasification 470 deg. C	"	Mixture of carbon & lime	-	62.1
Calcination/Regeneration Cyclone	"	Mixture of carbon & lime	-	100

In this case there is a discrete mixture of particles with a maximum size around 1000 μ. Finally, Figure 10 illustrates the type of particles collected during kerosene combustion at 870 deg. C of the sulphided, aged, Denbighshire bed. Here, there is a discrete mixture of particles which are mainly in the size range 50-1000 μ.

These results indicate an approximate correlation between fines product rate and particle size of the fines collected by the cyclone. The large cyclone particles were collected at the lowest production rate i.e. during kerosene combustion of sulphided, aged, Denbighshire stone. Also it would appear that kerosene combustion at 870 deg. C in BCR 1691 which gave the highest fines production rate resulted in the smallest particles being collected in the cyclone.

It is not possible, however, to extend this argument and correlate fines production rate with particle size of total fines produced because of differences in cyclone efficiency. One of the lowest cyclone efficiencies was recorded for the large cyclone particle sizes and one of the highest for the smallest cyclone particle size. If particle size of cyclone material was a true reflection of particle size of fines produced by the bed, then one would have expected cyclone efficiencies to have been highest with the largest particle size. Why cyclone efficiency should act in this way is indeed puzzling. However, it does mean that this information of fines size and appearance cannot be used with respect to fines production mechanisms.

Further tests on the fines collected showed that those collected during gasification flowed best and equally well for all three stones. Under all other conditions those from BCR 1359 and Denbighshire flowed more freely than those from BCR 1691. The superior flow characteristics of fines from gasification are attributed to their high carbon content which could be as much as 45% by weight. The superior flow characteristics of fines from Denbighshire and BCR 1359 stones in relation to BCR 1691 under all the other CAFB conditions are attributed to the presence of fewer of the very fine particles. Although these differences were encountered in the flow properties of the fines, we never encountered batch unit fines with the severe stickiness of pilot plant fines made during Run 4 startup with BCR 1691.

Batch unit fines and bed samples from BCR 1691 were analysed for calcium and silicon. The results showed that the batch unit conditions also caused the preferential loss of calcium from the bed observed in pilot plant Run 4. It appeared to begin during calcination and continue through the tests. As in the pilot plant, calcium lost from the bed did not appear in recovered fines but was lost from the system. The extent of calcium loss in the batch unit tests was not as great as found in the pilot plant where the SiO_2/CaO ratio increased to 0.41 compared with an initial value of 0.27. In batch unit cycle tests, the SiO_2/CaO ratio lined out at approximately 0.33.

During each test unit pressures were monitored to determine if any blockages were occurring downstream of the bed and after each test exit gas lines were dismantled and examined. Only one blockage of any significance was ever encountered. This occurred with BCR 1691 under kerosene combustion conditions at 870 deg. C and was traced to a period of cyclone rapper malfunction. This emphasises the importance of fines concentration in the gas stream in relation to blockages and shows that even with BCR 1691's high rate of sticky fines production, proper draining of cyclones with the aid of rappers where necessary prevents blockages.

SECTION VIII

RUN 5 PILOT PLANT MODIFICATIONS

MODIFICATIONS AFTER RUN 4 AND BATCH STUDIES

Between Runs 4 and 5 a number of modifications were made to the pilot plant to permit improved operations under conditions like those encountered in Run 4. The experience of Run 4 and data gained in the batch unit programme revealed that operation is more difficult with limestone BCR 1691 than with Denbighshire stone. The batch work indicated that BCR 1359, another high purity stone, should behave more like Denbighshire stone. However there are many locations where high purity limestone will be considerably more expensive than lower purity stones available locally. It therefore was desirable to assess more fully the consequences of operating with a lower purity stone. Although BCR 1691 is not necessarily typical of all low purity stone, it has deficiencies which, if overcome, would assure that the CAFB gasifier could operate with stones of a wide quality range. Five major changes were made to assist operations with the dusty and more easily agglomerated stone.

- External cyclone drainage
- Non obstructing regenerator pressure control system
- Flue gas recycle scrubber
- Regenerator overtemperature quench
- Flue gas stack scrubber

The revised pilot plant flow plan in Figure 11 shows these modifications. Other minor changes were made to the unit to improve pilot flame stability and to assist in diagnosing boiler performance.

Cyclone External Drain System

Pressure balance calculations on the gasifier cyclone return system indicated that there would be insufficient height of leg available to return fines to the gasifier through the internal passages if bed depth were increased to the levels desired for high sulphur recovery with low replacement rates of BCR 1691 stone. This problem increases in severity when fouling increases the pressure drop across the cyclone inlet. By using external cyclone drains, the pressure at the cyclone drains could be made independent of the gasifier bed pressure. It was not sufficient however just to drain the cyclones externally. With deep beds the rate of entrainment into the cyclones could be high and stone losses severe unless the coarse fraction were returned to the gasifier. Therefore an external system to both drain the cyclones and return the fines was needed.

Cooling Tower
Soft Water
Heat Exchanger
To Pressurization Unit
Air
Stack
Boiler
Fuel Oil
Air
Flue Gas Recycle
Air
Air
Air

After consideration of several designs, a system was selected which met the constraints of available space, pressure, and gas consumption. Details of the system appear in Appendix C. In summary the system for each cyclone consists of a conical bottom pot receiver mounted beneath each cyclone drain, a butterfly valve to isolate the pot from the cyclone when the pot is being emptied and a Warren Springs Laboratory pulsed flow powder pump to transfer solids from the conical pot to an overhead receiver, common to both cyclones. An elutriator removes the very fine fraction from the cyclone solids, and a pneumatic injector returns the larger size fraction to the gasifier bed. Nitrogen is the operating gas for the transfer system. Most of the time the butterfly valve beneath the cyclone remains open draining solids to the conical pot. At timed intervals the butterfly valve closes and the pot is pumped out to the overhead receiver.

Regenerator Pressure Control

To avoid a recurrition of the regenerator off gas line blockage which terminated Run 4, the pressure control valve was removed from the outlet line. This valve had become blocked by fine solids in the gas stream during that run. In order to vary regenerator back pressure, a new blower was installed to inject air into the outlet line just downstream of the cyclone outlet. Air flow from this blower is regulated by a control loop which senses the difference between gasifier and regenerator pressure and adjusts a valve in the air line to achieve the desired pressure difference.

Flue Gas Recycle Scrubber

Run 4 demonstrated that simple cyclones were unable to provide sufficient cleaning of the recycle flue gas stream to prevent gradual blockage of the gasifier air distributor nozzles. A venturi scrubber system was designed to provide greater cleanup; the scrubber was designed to handle 200 CFM of gas at a pressure drop of 14" w.g. Water is sprayed into the gas at the throat of a venturi. A knockout vessel at the venturi outlet removes the water and entrained dust. The venturi was placed on the suction side of the recycle blower to protect the blower from dust, and a recycle line was provided to permit a high gas circulation rate through the venturi even at low rates of flue gas flow to the gasifier.

Regenerator Quench

The circulation of fresh solids from the gasifier to the regenerator controls regenerator temperature. Upsets in the pressure balance between gasifier and regenerator or temporary obstructions in one of the solids transfer lines can sometimes interrupt this solids circulation and allow regenerator temperature to increase.

If regenerator temperature gets too high there is danger of sintering the lime particles and forming agglomerates. An emergency quench system was installed to prevent this occurrence. The lower regenerator bed thermocouple was connected to a controller which admits a flow of nitrogen to the intake of the regenerator air blower when bed temperature reaches the alarm point. The alarm was set to operate at 1100 deg. C. Nitrogen fed to the blower dilutes the regenerator air supply and reduces the rate of oxidation to prevent over temperature. The circuit is fitted with a manual switch so that the process operator can inject nitrogen at will in the event of other forms of upset.

Stack Top Gas Scrubber

To avoid particulate emissions to the atmosphere during periods of high lime losses from the gasifier cyclones, a final stage of water scrubbing was added to the pilot plant flue gas stack. Experience in Run 4 had indicated that the flue gas cyclone was not completely effective in recovering lime fines produced from HCR 1691 under combustion conditions.

The new scrubber consists of a section of ductwork shaped like an inverted "U" mounted on top of the stack. The down leg directs the gases into the top of a funnel shaped receiver which causes another reversal of gas direction upward to the atmosphere. Water is sprayed into the down leg and collected by the funnel. This water which picks up limedust by passage through the flue gas is conducted to a ground level settling vessel. Overflow from this vessel is circulated back to the scrubber nozzle by a centrifugal pump. The system is designed to circulate water to the scrubber at a rate of approximately 100 gallons/min.

SECTION IX

PILOT PLANT OPERATION - RUN 5

GENERAL

Objectives of Run 5 were to measure sulphur removal efficiency at different bed levels and lime replacement rates, to test feasibility of desulphurising gasification at temperatures above 930 deg. C approaching adiabatic conditions, and to determine effectiveness of pilot plant modifications in solving problems met in Run 4.

The test programme was to start up with precalcined lime from Denbighshire stone, operate for a test period with Denbighshire stone, and then switch to BCR 1691 stone. Provision was allowed for returning to Denbighshire stone if the BCR 1691 proved inoperable.

In actual fact, Denbighshire stone was used during the first two days and the final week of operation. BCR 1691 was used for the rest of the run.

Table 11 summarises the various periods of operation during the run. For purposes of computer identification of data, each run hour is designated by a decimal number with its whole part signifying run day and its fractional part the hour. February 6 was run day 1.

For example, 4.1630 represents 4.30 p.m. on February 9. The same system is employed on the abscissa representing time in the data graphs of Appendix G. These graphs show the variation with time of major operating variables during the run. Tables of detailed operating data also appear in Appendix G.

OPERATIONS SUMMARY

Operability of the pilot plant was much better in Run 5 than in any previous test. Prolonged test periods were achieved with lined out conditions. Some initial difficulties were experienced with some of the new equipment, and some of the problems encountered in earlier runs reappeared but in less severe form.

Heat up began on 3 February 1971 and proceeded smoothly without interruption. Precalcined lime from Denbighshire stone was used for the initial bed, and no difficulty was experienced in achieving fluidisation. During the first 108 hours of continuous gasification, four data periods were obtained, two with Denbighshire stone and two with BCR 1691. This first gasification period was terminated when pressure drop through the boiler became excessive. The increase in pressure drop was caused by a build-up of deposits at the entrance to the first pass of small boiler tubes. The gasifier was sulphated and decoked by controlled oxidation, and the boiler tubes were cleaned.

Table 13

Summary of Run 5 Operating Periods

Dates	Run Days	Limestone Feed	Number of Test Periods	Gasification Hours	Cause of Termination
Feb 6 - Feb 9	1.2083-4.1528	Dolomite/Lime	2		Pressure drop through boiler.
Feb 9 - Feb 11	4.1528-6.1389	BCR 1691	2	104	Change to BCR 1691 caused by interruption of gasification.
Feb 13 - Feb 14	8.1528-9.1639	BCR 1691	0	11	Regenerator Defluidization
Feb 16 - Feb 18	11.1215-13.1000	BCR 1691	4	45	Fuel Injector Failure
Feb 21 - Feb 22	16.1639-17.2208	BCR 1691	2	40	Regenerator Defluidization
Feb 25 - Feb 27	20.1806-22.1906	Dolomite/Lime	4	49	Cyclone Inlet Pressure Drop
Feb 28 - March 3	23.1300-26.1900	Dolomite/Lime	5	78	Voluntary - End of Run

Resumption of gasification was delayed by malfunction of the fines return system and by behaviour which suggested an obstruction in the regenerator. However removal of the regenerator distributor revealed no blockage. The trouble with the fines system was caused by chunks of coke dislodged during decoking.

Gasification resumed on 13 February but was interrupted 11 hours later by regenerator defluidisation and a regenerator hot spot of 1150 deg. C. Removal of the regenerator distributor revealed some agglomerated solids in the hot spot region. Cause of the defluidisation is suspected to be partial gas by-passing and lack of fines in the regenerator.

The next gasification period began on February 15 and lasted 45 hours. Five data periods were obtained with BCR 1691 stone. This period was interrupted by failure of one of the fuel injectors. The tip of the injector was destroyed causing poor local fuel distribution and local defluidisation. After sulphation and decoking, the other injectors were inspected and found to be intact. Since the damaged injector was nearest the start-up burner, it is possible that it had been weakened by over-temperature during the heat up period. The fuel injector was replaced, the boiler tubes were cleaned, and gasification resumed on 21 February. This gasification period lasted 40 hours with BCR 1691 stone and provided two data sets.

This period also was terminated because of partial defluidisation in the regenerator.

The regenerator superficial gas velocity, based on air and nitrogen inputs, was over 5 ft/sec during this period, more than enough for good fluidisation. The cause of the loss of fluidisation was again suspected to be gas by-passing, but this hypothesis was not confirmed. The difficulty was aggravated by malfunction of one of the regenerator blower-stages which limited the available air capacity. Following this interruption, the regenerator distributor and interior were again inspected.

The regenerator interior contained only a thin collar of fused material just above the distributor, not enough to explain the regenerator difficulties. The refractory lining of the distributor was rebuilt, and cracks in the lower portion of the regenerator were patched with fire clay before reassembly.

The next gasification period consisted of 49 hours with Denbighshire stone. Fuel rate was increased to test gasifier operation at 950 deg. C, with lower air/fuel ratio and reduced flue gas recycle. Pressure drop through the boiler prevented increasing fuel rate enough for completely adiabatic conditions but smooth operation was maintained at over 400 lb/hr oil rate and 19 to 20% of stoichiometric air-fuel ratio. Two test periods were obtained at the high fuel rate and two more at an oil rate of about 310 lb/hr.

No difficulties were experienced with regenerator defluidisation during this period. Gasification was interrupted for decoking due to rising pressure drop through the gasifier cyclones. The boiler was not cleaned during this down period. Gasification was resumed for a final 78 hours of gasification with Denbighshire stone. Six more data sets were obtained in this final period after which the unit was shutdown. Mechanical operation was smooth in the final period. The external gasifier fines return system functioned well, except that cyclone efficiency declined, and a reducing amount of fines was caught for return.

At the end of gasification, the unit was shut down without sulphation or decoking in order to permit examination of the interior with deposits in place.

EQUIPMENT PERFORMANCE

Equipment performance was greatly improved over that experienced in previous runs, particularly that of Run 4. A number of the new features operated well, but some continued to be troublesome throughout the run.

Stone Feeder

The stone feed system which used a vibrator in a pressurised shell to feed from a weighed hopper proved reliable throughout the run. Stone feed rates were usually quite steady and easily measured.

Regenerator Drain Valve

The gasifier bed level control system which used a pressure switch in the gasifier to activate a drain valve in the regenerator proved to be reliable and to give good control of gasifier bed depth.

Regenerator Pressure Control

No blockages were encountered in the regenerator off gas line during Run 5. This line and its control system had plugged continually during Run 4 start up. In Run 5 the line remained clear and showed no sign of pressure build up. The system used controlled introduction of excess air into the outlet line downstream of the cyclone and avoided restrictions in this line. It was not possible to operate the system in automatic mode due to the large pressure pulses introduced by the solids circulating system, but manual control of the pneumatic valve position proved satisfactory for control of the pressure difference between regenerator and gasifier. Pressure difference was regulated to within one or two inches water gauge. Normally the regenerator pressure was adjusted to be 1 to 5 in. w.g. below gasifier gas space pressure although higher differences were sometimes used.

Cleanliness of the regenerator gas line was also aided by continuous use of a pneumatic rapper on the regenerator gas cyclone. This rapper ensured drainage of solids from the cyclone walls. Also, the conditions that produced the very sticky fines, kerosene combustion in a bed of BCR 1691 stone, were avoided as much as possible.

Flue Gas Recycle Scrubber

The venturi scrubber on the flue gas recycle stream removed a great deal of lime fines from the gas, but was not completely effective. Some particles passed the scrubber, and some fouling of the recycle gas line, control valve, and gasifier distributor was encountered. The rate of gasifier distributor pressure rise in Run 5 was much less than in prior runs. In the initial stages of Run 5 there was frequent plugging of the inlet of the venturi throat itself with lime deposits. Increasing the gas flow through the venturi to the maximum rate available (estimated at 200 CFM) by using maximum recirculation eliminated plugging at this point.

The drain line from the water separator occasionally blocked and required cleaning.

On at least three occasions blockage of this discharge caused water carryover to the gasifier plenum itself with a consequent sharp decrease in gasifier temperatures.

Regenerator Over Temperature Protection

The new regenerator emergency quench system dilutes the inlet air with nitrogen when regenerator temperature reaches the set point of 1100 deg. C. This system proved to be quite valuable and avoided excess temperature several times when malfunction of the solids circulation system reduced lime flow rate through the regenerator. In only one case did regenerator temperature seriously exceed 1100 deg. C, and that was due to emptying of the quench N2 supply bottle before normal conditions were restored. This quench system is believed to be responsible for avoiding regenerator blockages by agglomerated solids which occurred in previous runs.

Boiler Pilot Flame

The new boiler pilot burner and its gas and air system were quite effective in providing a stable and reliable pilot. No difficulty was met in lighting the pilot over a wide range of conditions nor in keeping it lit.

Stack Top Washer

The water spray scrubber installed to prevent discharge of dust to the atmosphere was operated during part of the run when it appeared that some dust was passing the external flue cyclone.

When operated the scrubber appeared to be effective in avoiding dust emissions. Because of the diffuse upward discharge of gas from the system, it was not possible to obtain a quantitative measure of the actual dust content.

Corrosion of the water recycle piping in this scrubber system was severe because of SO2 from the regenerator which was remixed with the boiler gas in the stack.

Cyclone Fines Return System

The gasifier cyclone fines return system operated well during most of the run in spite of several deficiencies. Operation was trouble-free for most of the first 109 hours of gasification and for the final 127 hours. During other periods there were a number of upsets. Two problems were encountered in the first period:

(1) Dust worked its way back to the pneumatic control system through a pressure measurement line and caused stoppage. This problem did not recur after installation of a fine filter and additional N_2 bleed in the pressure line.

(2) Residual pressure remained in the conical dust receiver after a transfer of solids when the butterfly valve to the cyclone drain opened to begin refilling, this pressure caused a surge of gas back up the cyclone leg and upset the cyclone operation. The result was a burst of fines into the boiler after each transfer operation. These puffs were evident in the peaks observed in boiler SO_2 emission. This difficulty was removed by installation of a delay device which allowed pressure to discharge down stream from the conical receiver before the valve to the cyclone could open.

Most of the problems met during the mid-run period were caused by chips, flakes, and chunks which found their way into the cyclones and transfer system after each temporary shutdown and decoking operation. It was necessary to disconnect vessels and lines on several occasions to remove these flakes and chunks. In other cases these solids prevented good operation of the butterfly valves. When the butterfly valves failed to seat properly before a transfer, N_2 gas again blew back through the cyclone and sent dust to the boiler.

Installation of a chunk trap in the slurrisator drain during the run improved operation a great deal. Installation of additional chunk traps in the conical vessels will be a further aid.

Efficiency of the cyclones themselves deteriorated during the run. In the initial period efficiency was fairly high, and only a small amount of very fine solids entered the boiler. In later stages of the run the efficiency deteriorated, and a considerable quantity of quite coarse solids entered the boiler.

Inspection of the cyclones at the end of the run revealed them to be nearly completely choked with a mixed deposit of lime and carbon in the annular space between walls and gas outlet tube. The steel liners which had been installed before Run 5 were severely burned and distorted. The silicon carbide gas outlet tubes were strong, smooth, and intact. It is evident that demonstration of an effective way to maintain cyclone efficiency must remain an important problem area of U-tube work.

In view of performance of the process, it is clear now that recycle of cyclones fines to the gasifier without a means to achieve their regeneration is not desirable. The fines, with their large surface area, pick up a considerable load of sulphur and make a number of cycles through the gasifier and cyclones without entering the regenerator. If any of these fines escape the cyclone, they enter the boiler and cause loss of sulphur removal efficiency. A means of providing preferential regeneration of the fines is a desirable process feature.

Regenerator Operation

Defluidisation of the regenerator bed occurred twice during gasification in Run 5 and once under combustion conditions. The cause of this behaviour has not been established, but gas by-passing in some manner is suspected. The effects of such by-passing would be aggravated by lack of fines in the regenerator solids which would increase minimum fluidisation velocity. Two of the ways in which by passing could occur are leakage of gas through the solids circulation passages and leakage through cracks. It is possible for air to enter a crack in the refractory near the bottom of the bed, travel upward through the crack, and return to the vessel higher up. A small crack was observed in the regenerator wall and patched during the run, but it did not appear large enough to account for the troubles observed.

The fact that the tendency to defluidise became more severe with time during the gasification periods involved suggests that it may have been related to another time dependent factor such as the increase in gas space pressure which took place as gasifier outlet passages gradually fouled. Leakage of air back into the gasifier-to-regenerator solids transfer line and up the unused portion of the cyclone fines return leg could follow such a course. Such leakage would have to pass into the cyclone past the steel sleeve insert which had been dropped into the cyclone leg as a seal. However, distortion of the steel sleeve by heat following gasifier decoking is a distinct possibility. On the other hand, such a loss of air near the regenerator bottom does not accord with the apparent low SO_2 concentration measured in the regenerator gas. Indeed, sulphur material balance considerations imply that the gas flow through the regenerator was higher than that supplied by the regenerator air blower. It is probable that further tests will be needed to establish the cause of this unusual regenerator behaviour.

Boiler Deposits

Boiler deposits found within the boiler were of two types:

(1) Loose accumulations of dust or coarser particles in the soot trap areas at the boiler ends.

(2) Agglomerated deposits formed from very fine particles which build up at the inlets to the first pass of small fire tubes.

The loose accumulations of particles do not appear to present a long term problem. They would be subject to easy removal by normal soot blowing techniques.

the agglomerated fines represent a potential problem area which requires additional study to define its severity in large scale equipment. Certainly it is an inconvenience in our pilot plant equipment. However it must be stressed that the fire tube boiler used for our pilot plant tests is in no way typical of a water tube power generation boiler and the problems we have experienced may be typical only of the particular boiler we are using. From the point of view of deposit buildup the pilot plant boiler is far from ideal. Deposits are very local in nature; being found only at the inlets to the first set of water cooled fire tubes, where the gases change direction by 180 deg. in a downward direction.

They did not form after the first few inches of tube length nor were they found to any significant degree on a test probe inserted radially into the gas stream at the end of the main fire tube as a simulation of a superheater tube in a water tube boiler. These deposits evidently form from fine particles which are in a sticky state following their passage through the flame. The growth of the deposits was faster during the first 109 hours of Run 5 gasification than during the final 127 hours. Whether this difference was due to the presence of BCR test stone in the first period or to the eroding effect of a higher concentration of coarse particles during the final period remains to be established. It is possible that deliberate injection of a small amount of coarse stone into the boiler could prevent deposit formation. No deposits were found at the tube inlets during the 111 hours of gasification in Run 2 during which a high concentration of solids passed through the boiler.

PROCESS PERFORMANCE

Table 14 lists values of operating conditions and results for the various test periods of Run 5. Each value is the average for four hours operation. In most test periods a set of solids samples was collected for analysis.

Sulphur Removal

The degree of sulphur removal in the pilot plant is calculated from the measured SO_2 and CO_2 contents of the boiler flue gas compared with sulphur and carbon contents of the boiler fuel. The carbon content of the pilot burner propane is considered in this calculation, as is the CO_2 released by calcination of limestone makeup.

During Run 5, sulphur removal efficiency (SRE) varied from 60 to 99%. Appendix figure D-1 shows the hourly levels of SRE along with other important variables. Figure 13 shows the effect of lime replacement rate on SRE for individual test periods. It is apparent that lime replacement has a major effect on SRE.

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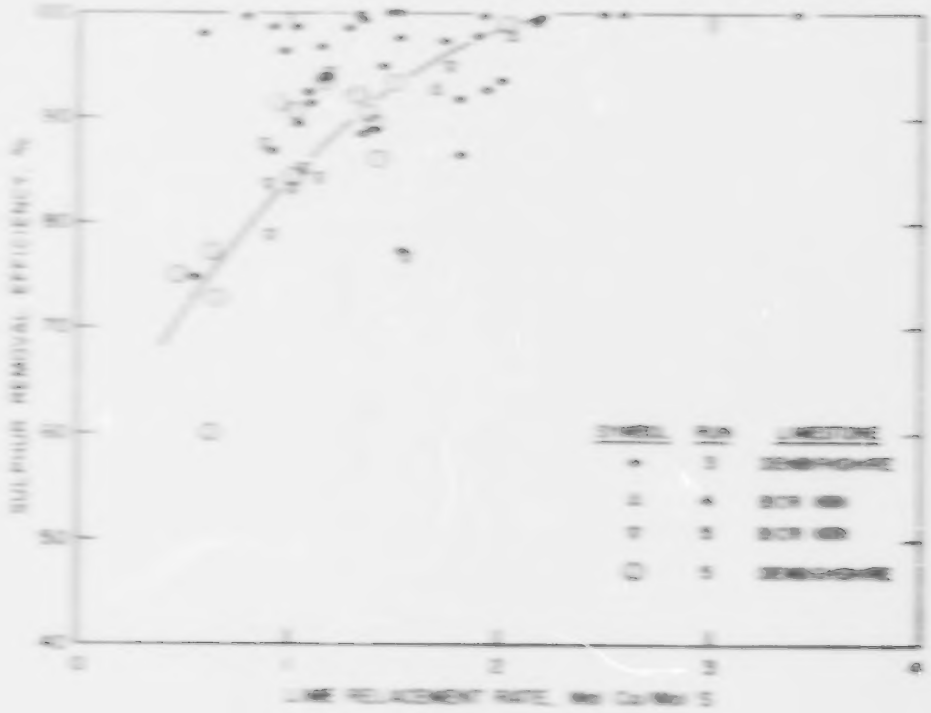
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3
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5

Figure 11. [illegible]

The effects of other variables are less clearly defined; in particular, increasing gasifier bed depth did not produce the expected improvement in SO2 over results obtained in earlier runs with shallower beds.

It appeared in this run that the beneficial effect of increased bed depth was offset by increasing fines loss rate. The loss of fines hurts sulphur removal efficiency in two ways. Firstly it removes the high surface area fraction of the lime which has greatest potential for sulphur pickup, and secondly, when sulphur laden fines enter the boiler they partly regenerate to contribute SO_2 to the flue gas.

The effect of bed depth was also obscured by the fact that deterioration in cyclone performance made it difficult to operate with deep beds at very low lime replacement rates.

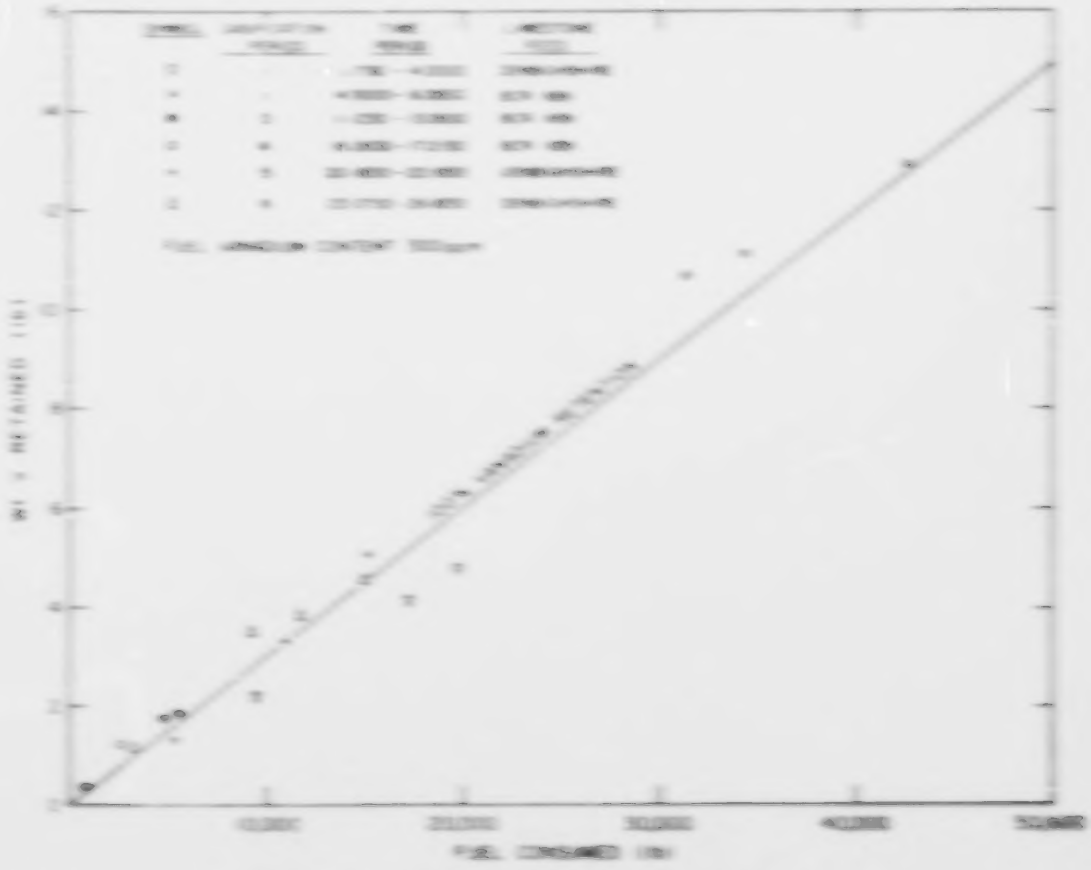
The test at 900 deg. C gasifier temperature on day 21 produced sulphur removal efficiency over 90% and demonstrated the feasibility of operating at this temperature with a low air/fuel ratio. The fuel rate in this test was 469 pounds per hour, the highest yet used in the pilot plant. Pressure drop in the boiler prevented increasing fuel rate still further to test completely adiabatic gasifier operation without flue gas recycle.

Metals Retention

Comparing the metals content of spent lime with the metals content of fresh limestone and fuel oil, it is possible to estimate the degree of metal retention by the solids. Figures 13, 14 and 15 compare the retained weights of vanadium, sodium, and nickel with quantities of these metals fed during various operating periods of the pilot plant.

Vanadium retention is essentially complete, in agreement with predictions based on batch unit tests. Sodium retention was 15%, somewhat higher than the 20% level obtained in batch tests. Nickel retention, which was not studied before, averaged 75%.

There appeared to be no significant difference between metals pick up efficiencies with Newfoundland and Nova Scotia stones; however because of differences in stone loss rates, there is a difference in absolute metals retention levels in the unit. With Newfoundland stone there was practically no lime loss from the system. That lime which escaped the gasifier cyclones was caught either in the boiler or by the external flue gas cyclones. However with Nova Scotia, losses amounted to as high as 40% of the lime replacement rate and of course any associated metals were lost as well. While not affecting the ability of the CAFB gasifier to remove vanadium as a source of high temperature corrosion of boiler superheater tubes, the loss of metals on lime particles would be a pollution factor which could be reduced even further by increasing the efficiency of the particulate removal portion of the system.



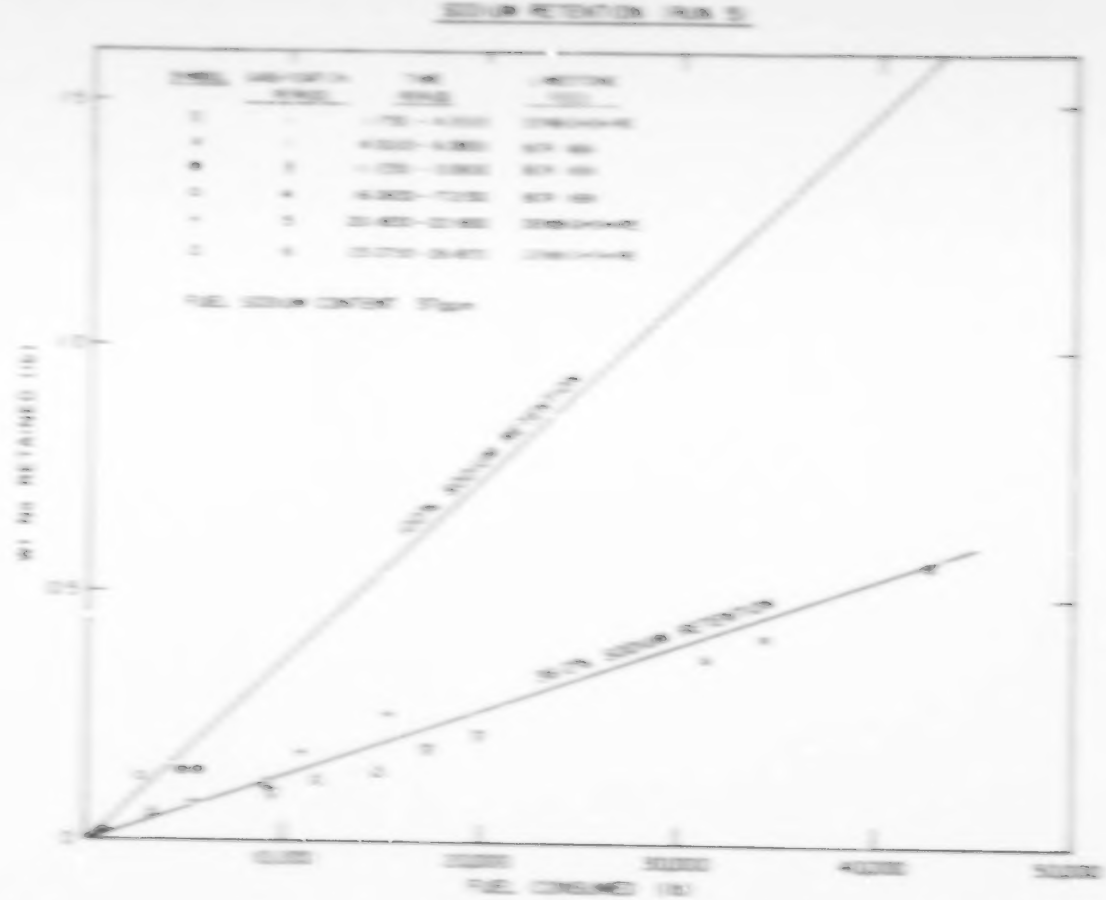
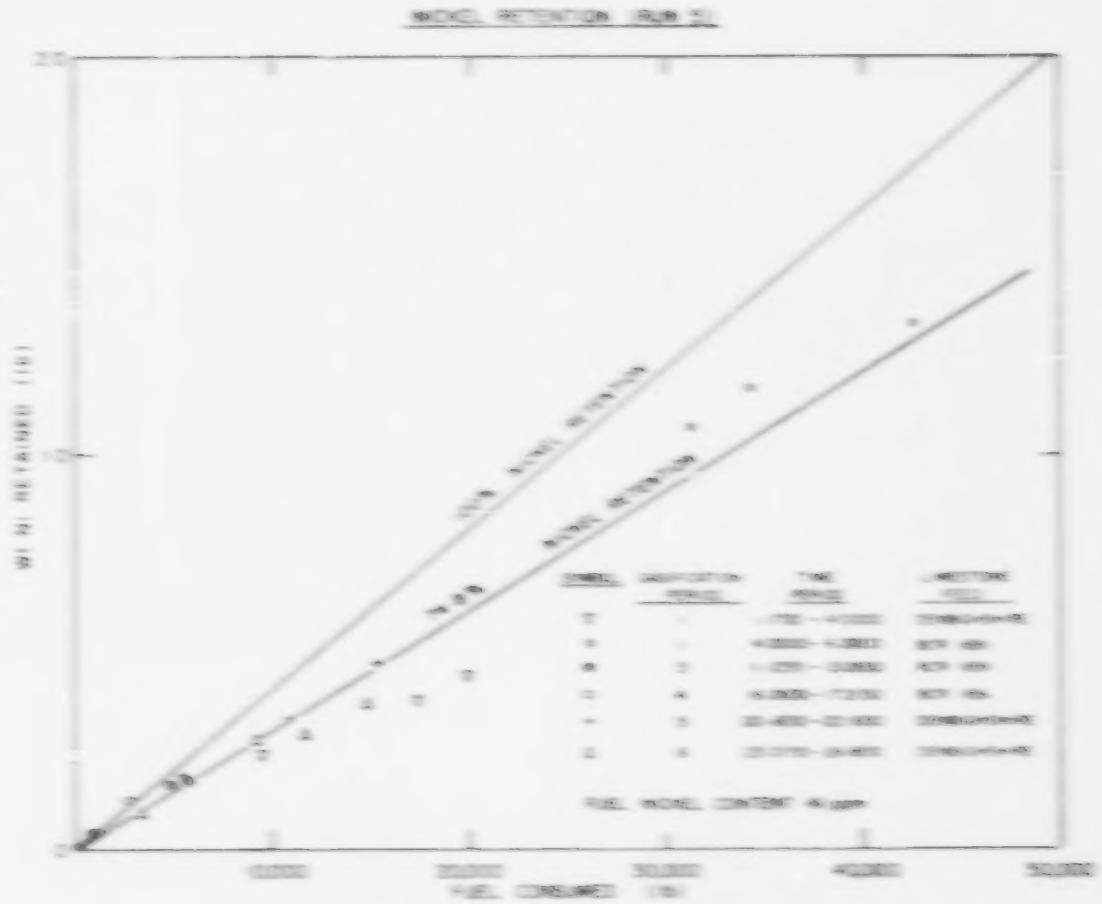


Figure 24 Sodium Retention (Run 1)



Solids Losses

To obtain a more comprehensive picture of solids losses during Run 5, both the amount of material emitted by the gasifier into outlet and the amount escaping the external cyclone into the stack have been computed.

Results are summarised in Table 15. The water scrubber on the stack top was able to catch the worst of the material which escaped the cyclone, but since the quantity recovered by the scrubber was not measured, it is included with the stack losses.

Table 15

Summary of Solids Loss - Run 5

Time Day:Hour	Limestone Feed	Make-up Rate (lb/hr)	Gasifier Fluid Bed Depth (inches)	Loss Rate (lb/hr) Gasifier	Stack	Bed SiO_2/ CaO^2 Ratio
1:1530	Denbighshire	7.3	27.5	0	0	0.009
1:2030	"	7.3	22.5	5.3	2.7	0.006
5:1030	BCR 1691	60.0	33.0	29.9	23.3	0.703
6:0730	"	27.5	26.7	18.3	15.3	0.336
12:1730	"	26.0	31.5	31.0	22.2	0.224
13:0330	"	26.0	27.5	21.5	14.8	0.217
17:1130	"	21.3	35.5	10.8	9.7	0.199
17:1730	"	41.0	37.3	33.2	25.7	0.185
21:1830	Denbighshire	22.0	30.7	18.3	12.0	0.051
22:0630	"	22.0	28.7	17.2	10.3	0.053
23:1730	"	17.0	30.0	13.2	6.3	0.038
25:0530	"	25.0	35.7	16.2	0.1	0.023
25:1430	"	25.0	36.4	17.0	4.3	0.016
26:0830	"	17.5	35.7	13.2	1.0	-
26:1730	"	17.5	39.2	12.0	0	-

Taking gasifier losses first, it is apparent that these varied considerably during the run. Previously, in batch units it had been found that gasifier loss rate was dependant on make-up rate and bed depth. (1) In Run 5, however, the situation was more complicated since cyclone performance was known to have deteriorated sometime during the run. This was evident from the after-run inspection which showed the cyclones to be in poor condition. Statistical analysis of gasifier loss rates showed an inconsistency between the first and second data points (1.15t/h and 1.20t/h). Also all later variations for both stones could be explained by changes in bed depth and make-up rate in a single correlation.

From this analysis, the following was deduced. Firstly, cyclone performance deteriorated between the 1.15t/h and 1.20t/h data points and did not alter appreciably thereafter. The reason for the deterioration is not as yet clear. Secondly, gasifier loss rates at constant cyclone performance were shown to depend on make-up rate and bed depth. Whilst the effect of make-up rate was similar to that observed in batch units, bed depth appeared to be less significant. Thirdly, gasifier loss rate was independent of limestone type in this instance. This would not always be the case. Here, it would appear that the cut-off point for the reduced performance cyclones and the attrition patterns for the two limestones is combined to cause this phenomenon.

For stack losses, a very different picture emerged when these results were examined in detail. The two limestones behaved differently. Under all conditions examined, stack losses were small when the bed was composed mainly of Denbighshire limestone. With predominantly BCR 169 (5.1t/h to 7.17t/h), however, they were appreciable. They were also higher when of the order of 20% of the bed as estimated from CO_2/CaO ratios was BCR 169 (21.8t/h to 22.6t/h). Statistical analysis also showed that stack losses from a BCR 169 bed correlated with make-up rate and bed depth.

Since the performance of the stack cyclone did not change during the run, these variations in stack losses can only be explained if it is accepted that BCR 169 produces a fraction of material of much smaller particle size than Denbighshire. This was been indicated from Run 4 and batch test data.

Since the stack cyclone was designed to have the same efficiency as the gasifier cyclones, the results from Run 5 indicate that a gasifier following the same principles for solids handling as the pilot plant and with gasifier cyclones fully operational would give negligible gasifier losses with a limestone of Denbighshire type and gasifier losses of the order of stack losses with a limestone of BCR 169 type.

Bed Homogeneity

Both Denbighshire and BCR 1691 limestones were tested in Run 5. A measure of bed homogeneity with respect to limestone type was obtained by analysing bed samples for silica and calcium oxide and comparing the ratio of these two compounds. The silica to calcium oxide ratio for the high purity Denbighshire stone is much lower at 0.008 than BCR 1691 at 0.28. Results are summarised in Table 16.

For the first two data times, the system was completely homogeneous since only Denbighshire stone had been added. During the subsequent BCR 1691 test period, the bed always had a SiO_2/CaO ratio below that of the raw stone. Also no persistent increase in the ratio was observed.

During the final Denbighshire test period, a reduction in SiO_2/CaO ratio took place and the ratio of the raw Denbighshire stone was approached. Material from the stack cyclone gave similar results to that from the regenerator cyclone throughout. During the BCR 1691 test period, the SiO_2/CaO ratio in these fines was generally higher than that for the raw stone.

The significance of the SiO_2 rating during the BCR 1691 tests is somewhat obscured by the fraction of Denbighshire stone which remained in the bed following changeover to BCR 1691.

We believe that the low SiO_2/CaO ratios in the bed during BCR 1691 tests was due to residual Denbighshire stone. The absence of any increase in ratio over the period can be attributed to the addition of Denbighshire stone during burnout and maintenance periods between 6.0730 and 12.1630 and also 13.0600 and 17.1130.

We discount the possibility of a preferential loss of silica from the beds being the cause of the low results even though higher SiO_2/CaO ratios were observed in cyclone fines. The reason being that beds composed only of BCR 1691 in batch tests and Run 4 showed the opposite effect in that silica was concentrated in the bed. In those instances the fines which were trapped also showed a higher ratio than the raw stone indicating that the material lost completely from the system was rich in calcium. Assuming the same to have happened here with the BCR 1691 fraction of the bed, then the higher SiO_2/CaO ratio in the fines can be explained by BCR 1691 being lost preferentially.

After the final change to Denbighshire feed, the SiO_2/CaO ratios indicated that some BCR 1691 was present in the bed to the end of Run 5, albeit in ever decreasing amounts.

Particle Size of Solids

The particle size distribution of solids in the reactor beds depends on size of the feed, particle attrition, and effectiveness of the cyclones in returning fines. Sieve analysis of a number of gasifier and regenerator samples are presented in Appendix E. Histograms of the average stone feed and two sets of gasifier and regenerator bed samples are given in Figure 16.

The two sets of bed samples illustrated represent extremes of the samples taken. Performance of the fines return system was poor at 11,040, and fines were being lost from the unit. The cyclones and fines return system were operating relatively well at 21,160 as shown by the larger fraction of small particles.

These figures show that the fraction of particles in the 250 to 1400 micron size range was increased in the unit at the expense of both larger and smaller sizes. The fraction below 250 microns in the gasifier was lost altogether while the quantity of material above 1400 microns was reduced.

Poor performance of the fines return system, as at 11,040, causes significant loss of particles as large as the 600-850 micron range. Regenerator and gasifier beds were quite similar in size distribution with the regenerator showing a slightly higher fraction of fines.

Figure 17 shows the variation, during the run, of the fraction of bed in the size range below 600 microns.

This figure indicates a rapid deterioration of fines return effectiveness during the early part of the run. It indicates that fines return was restored during decoking before day 21 startup but again declined toward the end of run.

A similar pattern is shown in Figure 18 where average particle size of the bed is plotted against run time. The particle size here is calculated from sieve analysis, using the relationship:

$$d_{sv} = \frac{\sum w}{\sum w/d}$$

which gives a surface area mean particle size.

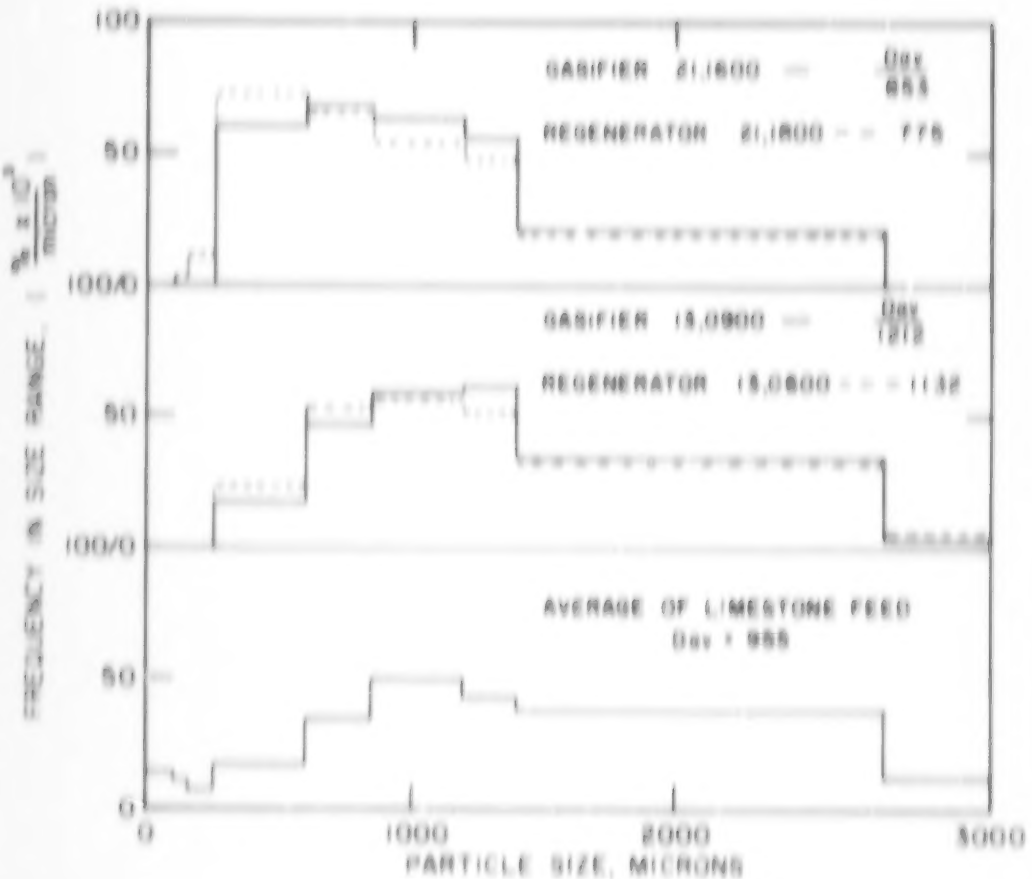


Figure 16 Size Distribution of Fluid Beds (Run 5)

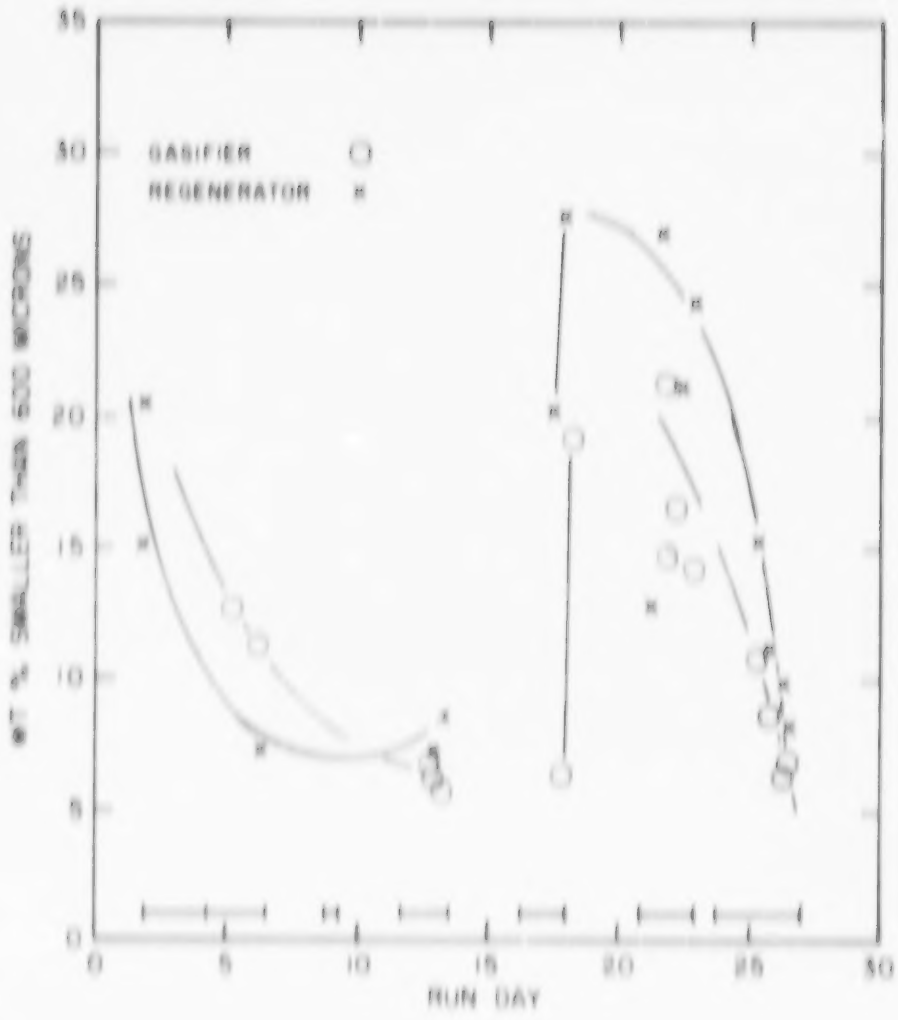


Figure 17. Gasifier and Regenerator Fines During Run Extension (Run 5)

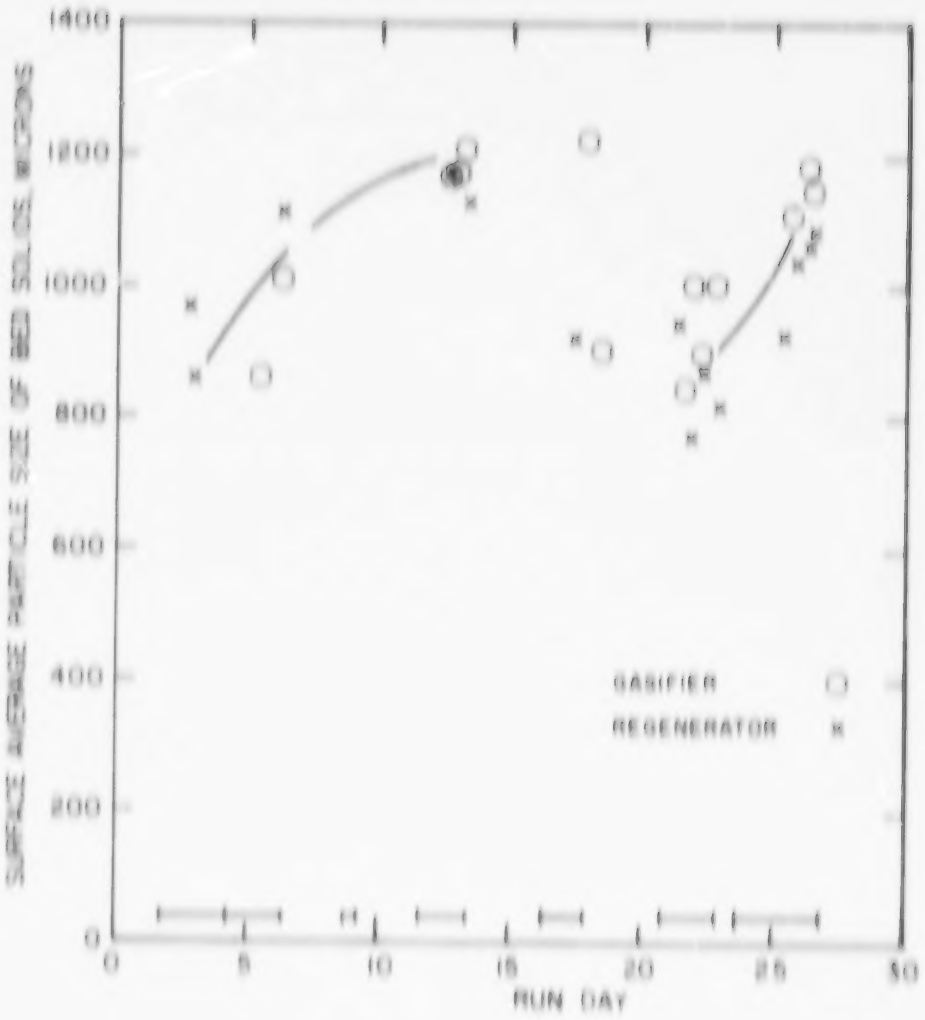


Figure 18 Average Size of Bed Particles (Run 5)

Nitrogen Oxides

The concentrations of nitrogen oxides measured in the boiler flue gas during Run 5 are compared in Table 17 with values measured in previous tests. All samples were taken from the boiler flue gas by means of gas sample bags and analysed off line in the laboratory. A chemiluminescence method was used for Run 5 samples.

The ASTM disod phenol-disulphonic acid method was used for the other samples. Laboratory cross checks have indicated that the two methods are in agreement.

Run 5 results agree generally with earlier gasification test results. All of the gasification results have lower NO_x concentrations than the tests with the oil burner. This improvement during CARB gasification is probably due to a combination of the effects of two stage combustion and the use of flue gas recycle. It is possible that nitrogen compounds in the fuel are converted to a harmless form during gasification, and that flue gas recycle reduces maximum flame temperature which reduces equilibrium NO_x concentration.

Even with the original oil burner operation the NO_x concentrations were low when compared with concentrations in the flue gas of large power station boilers. This difference in overall level is believed to be caused by the close proximity of the flame to the large water cooled surface in the fire tube boiler used in the pilot plant. Therefore, although the reduction in NO_x level caused by CARB gasification is believed to be realistic, the absolute low level achieved probably would not be reached in large power generation boilers where flame temperatures are much higher.

Thermal Behaviour

Most of the heat released by partial combustion of fuel in the gasifier is retained as sensible heat in the gas going to the burner. The heat release in the gasifier has been estimated from thermal equations for the masses and heat contents of the various streams entering and leaving the gasifier. Table 18 lists results for Run 5 test conditions. The equations used are explained in Appendix 5.

Heat losses from the pilot plant gasifier bed, based on reasonable values of thermal conductivity and heat transfer coefficients, amount to about 1% of the heat released in the gasifier. Depending on time replacement rate, about 2 to 4% of the heat released goes to calcine stone and raise the lime to gasifier temperature. At the air/fuel ratio employed, which did not deviate much from that of stoichiometric, the heat release per pound of fuel was estimated to be approximately 1100 Btu/lb corresponding to 151,000 Btu/lb mole of oxygen.

Table 19

Heat Release in 2025 Cessfire

Tab 1

Run Time	Run Time % of Total Simulation	Heat Lost % of Heat Release	Calibration and Other Ventures % of Heat Release	Heat Trans Implementation % of Heat Release	Cessfire Heat Release (MMBtu per Day)	Oil Heat Release (MMBtu/D)	L. Heat Release (MMBtu/D)
1.1130	30.8	.71	1.49	4.33	1.26	1172.	134659.
1.1450	30.4	.78	1.74	5.13	1.33	1165.	151409.
1.1530	30.4	.71	1.36	5.14	1.26	1167.	134461.
1.1210	30.1	.66	1.49	4.44	1.29	1145.	134062.
1.1040	30.1	.61	4.42	4.88	1.31	1176.	153768.
1.1770	30.4	.77	1.51	4.71	1.34	1174.	137925.
1.1210	30.4	.74	1.56	5.44	1.33	1165.	149531.
1.1940	30.6	.97	1.91	7.17	1.39	1225.	145776.
1.1030	30.7	.61	1.40	5.71	1.26	1115.	134344.
1.1860	30.6	.86	1.71	5.79	1.26	1165.	152362.
1.1230	30.3	1.24	1.1.	5.33	1.35	1138.	145641.
1.1730	30.9	1.24	4.33	5.54	1.28	1186.	144315.
1.1940	30.7	.81	1.33	4.76	1.39	1247.	149426.
1.1890	30.3	.81	1.47	4.27	1.40	1277.	1356[illegible].
1.1730	30.6	.86	1.61	1.11	1.27	1164.	155843.
1.1890	30.8	.88	1.59	5.14	1.26	1178.	154177.
1.1730	30.2	.97	4.92	5.30	1.29	1145.	162922.
1.1610	30.1	1.34	1.44	1.89	1.27	1131.	162062.
1.1530	30.1	1.1.	1.40	5.42	1.25	1084.	160495.
1.1930	30.9	1.17	1.43	6.28	1.23	1148.	160115.
1.1890	30.7	1.16	1.57	5.74	1.25	1144.	161551.
1.1630	30.8	.80	1.33	5.77	1.26	1225.	1[illegible].
1.1230	30.1	1.23	1.37	1.98	1.24	1124.	144185.

Figure 13 shows the observed variation of fuel heat release with air/fuel ratio, and compares the measured values with a line calculated for a release of 155,000 BTU/mole of oxygen. Values from all the pilot plant runs to date fall along this line. This value of 155,000 BTU/mole O_2 agrees well with heat release calculated from the fraction of carbon and hydrogen oxidized and the CO/CO_2 ratio formed in the gasifier. Details of this calculation also appear in Appendix E.

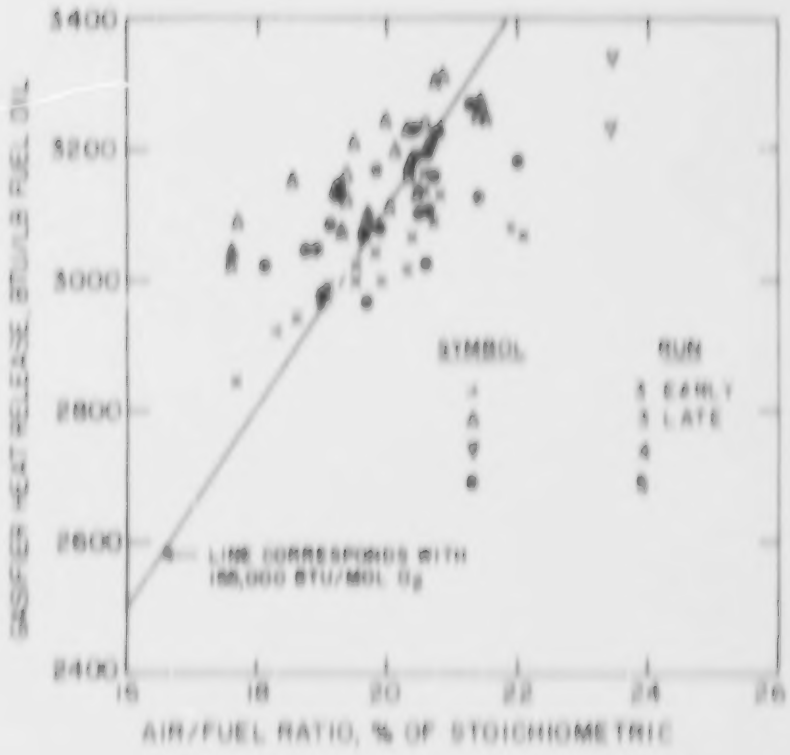


Figure 13 Heat Release Vs. Air/Fuel Ratio During Gasification

Product gas composition

Four samples of the gasifier product vapour were collected during the run and analysed by gas chromatograph. This analysis gives dry gas composition on a water and liquid hydrocarbon free basis. Results are listed in Table 19.

Table 19.
Product gas composition

Run 3

Sample time	22:10	22:17	23:40	24:10
Composition, (vol % dry & liquid free basis)				
N_2	51.7	52.5	53.8	54.1
CO_2	10.82	10.89	10.88	10.17
CO	8.16	7.58	9.36	8.93
H_2	6.86	7.75	6.42	5.80
CH_4	6.75	6.29	6.09	6.04
C_2H_4	5.36	4.84	3.81	4.18
C_2H_6	-	.11	-	-

A material balance calculation on the product gas composition and unit feed rates permits an estimate of the quantity and composition of that portion of the product vapour which is missed by the gas chromatograph because of condensation in lines or in the sample container.

In such a calculation nitrogen and oxygen balances are forced to zero and the amount of hydrogen and carbon not accounted for is assumed to make up the liquid fraction. The flue gas recycle stream was assumed to contain the same H_2/CO ratio as the boiler products of combustion. Table 20 summarizes results of this calculation. Details appear in Appendix Table F-1.

The results are fairly consistent in indicating the fractions of carbon and hydrogen oxidized and the quantity of carbon which goes either to coke or heavy hydrocarbons. The results on hydrogen disappearance show greater variability however.

This variability is probably due to the method of calculation which depends on finding small differences between relatively large numbers.

In particular, the hydrogen/carbon ratios of .85 and .91 calculated for the heavy components on day 21 appear to be unreasonably low; it is unlikely that the true H/C ratios could be much less than 1/1. These results show the desirability of obtaining accurate samples of the total gasifier product, including light and heavy components. However, collection of such a sample is quite difficult in practice, and will require development of a suitable quench and recycle system to collect the liquid fraction without plugging.

Regenerator

Performance of the regenerator during run 5 appeared to be less satisfactory than in earlier tests and was somewhat inconsistent. It was disappointing in that sulphur concentration in the off gas and selectivity of calcium sulphide oxidation to calcium oxide plus SO_2 appeared to be much lower than the levels which earlier runs had shown to be possible. Results were inconsistent in that the apparent sulphur production rate of the regenerator could account for only about half the sulphur being absorbed in the gasifier.

Furthermore, SO_2 ratios based on gas analysis did not agree with SO_2 based on sulfite analysis. Obviously this latter requires additional study to locate the cause of the discrepancy.

The run data shown in Appendix Figure D-2 are the gas analysis based values of SO_2 concentration and selectivity.

Table 20

Summary of Gasifier Component Distributions
(Calculated from Flows and Product Gas Composition)

Sample Time	22:10(C)	22:17(C)	26:04(C)	26:18(C)
Oxygen In, % of Total				
With Air	57.7	59.4	55.9	55.4
With Flue Gas Components	33.7	33.6	33.0	33.6
From Solids Reactions	8.6	7.0	10.0	10.0
Oxygen Out, % of Total				
As Oxygen in Ash	1.7	1.8	1.3	1.5
As Carbon Oxides	63.7	61.8	65.4	71.9
As H_2O (by Difference)	34.5	36.7	34.4	26.5
Hydrogen In, % of Total				
In Oil Feed	94.1	94.1	94.1	94.1
In Flue Gas H_2O	5.9	5.9	5.7	5.7
Hydrogen Out, % of Total				
As Dry Gas Components	84.4	70.5	87.1	89.9
As H_2O	11.0	13.0	12.4	10.7
As Heavy Components (By Difference)	4.0	4.5	10.5	17.4
Carbon To Gasifier, % of Total				
In Oil Feed	93.8	93.5	93.9	93.1
In Flue Gas, C oxides	5.8	5.6	6.0	6.1
In Ocean	1.0	0.9	1.0	1.1
Carbon From Gasifier, % of Total				
As Oxides from Gasifier	40.8	38.7	41.2	37.1
As CO_2 from Regenerator	1.0	0.8	0.8	0.7
As Hydrocarbon in Dry Gas	37.0	33.5	48.1	46.4
As Heavy Components (By Difference)	21.3	27.0	10.3	16.0
C Oxidized; % of Feed	15.8	14.9	15.9	14.5
H Oxidized; % of Feed	14.6	12.9	14.8	14.5
C in Heavy Components; % of Feed	8.3	10.0	3.1	5.1
H in Heavy Components; % of Feed	4.3	4.8	20.4	11.1
CO/CO_2 in Fresh Oxides	1.104	.97	1.44	1.45
H/C in Heavy Components	1.89	1.24	1.84	1.41
Air/Fuel Ratio, % of Stoichiometric	9.5	10.8	10.1	10.1

Table 12

Summary of Regression Performance

Run 3

Time	Gas Flow	Temperature	Volumetric Air to Coal Gas Analysis	Volumetric Air to Coal Oxygen Analysis	Proximate Analysis Dry Coal Moisture %	Proximate Analysis Dry Coal % of Coal	Ultimate Analysis By Difference Moisture	Ultimate Analysis By Difference % of Coal	Coal Flow Rate lbs/hr
1:15:01	2.84	1.42	27.8	50.1	2.87	30.8	7.55	54.4	72.8
1:30:01	2.72	1.41	27.4	52.4	2.82	29.7	7.44	56.4	40.3
1:45:01	2.65	1.39	40.4	70.3	4.28	44.3	7.83	40.1	43.8
2:00:01	2.72	1.44	40.3	41.3	2.67	28.9	7.83	43.1	46.4
2:15:01	2.88	1.47	30.4	54.2	3.11	34.5	7.41	77.1	84.4
2:30:01	2.87	1.82	32.4	53.5	3.83	40.7	6.68	70.2	94.1
2:45:01	2.76	2.04	37.1	40.7	4.13	47.8	7.04	53.4	87.9
3:00:01	2.73	1.83	34.2	40.7	4.21	45.2	7.13	54.8	84.2
3:15:01	3.01	2.37	44.2	74.1	4.99	53.7	9.62	84.1	93.1
3:30:01	2.69	1.78	28.2	50.1	3.01	26.9	5.77	50.9	92.9
3:45:01	2.85	1.71	30.2	64.1	4.24	40.7	6.21	54.9	87.2
4:00:01	2.90	2.01	32.7	64.3	3.32	34.1	6.83	49.6	90.4
4:15:01	2.91	1.91	41.1	54.2	4.14	40.9	5.94	54.9	85.2
4:30:01	2.74	1.88	32.1	43.7	3.04	27.4	5.24	52.3	91.7
4:45:01	2.92	1.72	34.8	53.3	2.69	27.3	5.80	54.1	90.1
5:00:01	2.74	2.17	27.1	34.5	3.14	32.8	4.07	40.2	89.9
5:15:01	2.71	1.96	34.1	44.8	3.91	39.2	5.28	53.1	90.8
5:30:01	2.87	1.94	31.7	70.6	3.34	34.1	7.82	58.9	87.4

Table 2: compares regenerator sulphur emission figures based on gas analysis with values calculated from solids analysis. Solids compositions were those of the gasifier and regenerator beds. It is apparent that sulphur production values based on the solids analysis are much higher than the gas analysis based figures. Similarly, the calculated values for oxidation selectivity are much higher when based on solids analysis. The solids based selectivities for [illegible]s run are in good agreement with the gas analysis based selectivities of Run 1 (figure 18 of Reference 1).

In Run 5 the level of sulphur in the bed was relatively low compared with the sulphur content of the cyclone fines. With the external cyclone fines return system these fines had little chance to be regenerated. In Run 1 however, fines from one cyclone drained to the regenerator. This arrangement must have decreased the sulphur content of the circulating fines and reduced the effect of fines loss to the boiler on sulphur removal efficiency.

SULPHUR BALANCE

Use of the gas analysis figures for regenerator sulphur output leads to very low sulphur material balances for the unit. Assuming that measured values of sulphur removal efficiency are correct, it is not possible to account for the missing sulphur by assuming that it left with the lime purge stream of fines losses. A fault in the regenerator off gas analyser could explain the discrepancy, but checks and calibrations made on the instrument during the run indicated that it was functioning properly. Similarly, a larger gas flow through the regenerator than measured would cause a low estimate of sulphur production.

Such a large error in gas measurement is unlikely because air, to the regenerator is measured both by orifice and gas meter, nitrogen to the solids transfer system is metered, and nitrogen to instrument bleeds is negligibly small. The only other possibility is a major leakage from gasifier to regenerator, and again this is believed unlikely. To help solve this mystery, additional analyses and measurements will be made in future runs: gas flow rate out of the regenerator will be checked by a helium tracer method, and a gas chromatograph will be used to check regenerator gas composition for SO_2 and other sulphur compounds.

We also plan to modify the boiler flue gas sampling system to reduce further the possibility of losing sulphur in the sample lines and filters. It is possible that SO_2 is absorbed on lime dust which enters the sample line, and such absorption would produce an optimistic estimate of sulphur removal efficiency. Such errors are believed to be small however, as the lines are

frequently cleaned, and spot checks with Draeger tubes (direct reading SO_2 colour change tubes) made directly through the boiler duct sample point agree with the continuous reading instrument.

SECTION X

SCOPING OF ENGINEERING EFFORT

GENERAL

As part of this project, Esso Engineering, Florham Park, New Jersey, USA, was requested to scope the engineering effort which might be required to carry CAFB from its present stage of development through the construction, startup, and testing of a large scale demonstration unit. A 100 MW scale unit was assumed as a basis. This scoping study is summarized here. Detailed results have been supplied to the Environmental Protection Agency in a separate memorandum.

SUMMARY

The total development of CAFB through a 100+ MW demonstration test period is expected to take about 6-½ years and require $1,120,000 in engineering effort. Of this total, $570,000 is for developmental engineering and pilot plant guidance; the remaining $2,750,000 is associated with the demonstration project. Optimistically, the total development might be accomplished in 4-½ years with $2,520,000 of engineering effort. Conversely, greater costs and times could be experienced.

The approximately 1 MW pilot plant at Abingdon can be scaled to 100+ MW without an intermediate pilot plant but with some risk. To reduce the risk, large scale mock-up studies of the fluidization system and special engineering development of critical equipment should be carried out. This scope includes the engineering manpower for these studies but does not include money to build any large scale test rigs. These are assumed to be included in the laboratory programme.

DISCUSSION

The approach used in preparing this study was to assess the state of development and the complexity of the process and then determine a reasonable schedule to carry the development to completion consistent with the criteria of reasonable risk. Previous experience in process and project developments similar to this was used as a guide. An estimate was then made of the various types of engineering activities required during the development and the extent of the effort required for each activity. This estimate excludes engineers who are directly associated on a full-time basis with research activities or the operation of pilot plants or demonstration plants.

The total programme has been divided into four activities, and a "most probable" and an "optimistic" development schedule has been estimated for each activity. Of course, there is also the possibility of a longer and more costly development programme if developmental problems prove extensive or if significant modifications are required to the demonstration plant due to start up difficulties.

The four activities in the development programme are:-

(1) Process Development
(2) Process Design
(3) Detailed Engineering, Procurement, Erection
(4) Startup and Test

The engineering effort is summarised in Table 22. The costs shown there are grouped to distinguish the engineering guidance costs during pilot plant development work from the engineering costs associated with design, erection, startup, and test of the large demonstration project. The project work is divided into two categories; basic engineering and the prime contractor's effort. Basic engineering includes the basic process design, owner's interest protection and follow-up during the detailed engineering and erection stage, and the engineering associated with start-up and testing. The prime contractor's effort involves the engineering required for mechanical design and erection of the demonstration plant.

In this schedule it was assumed that a client for the demonstration plant would be obtained near the end of the small scale development phase at which point a site would be selected. Certainly the earlier the client is located, the sooner it will be possible to direct both engineering and pilot plant activities towards a specific project with improved chances for shortening the time and cost of the development work.

The schedule and cost of the development activity is a major uncertainty in this type of effort. The schedule depends to a large extent on the degree of effort expended and the degree of risk which might be considered acceptable when starting the demonstration plant design. The estimate of 18 months for additional small scale development assumes minimum future pilot plant problems, very little process optimization, and a higher risk in proceeding with the demonstration plant than if more extensive (experimentation and engineering) development work were undertaken.

TABLE 22

CAFR DEVELOPMENT PROGRAMME
SUMMARY OF ENGINEERING EFFORT(1)

	Most Likely		Optimistic	
	MH (2)	Dates	MH (2)	Dates
DEVELOPMENT	570	1/73-7/75	360	1/73-7/74
PROJECT				
• Basic Engineering				
- Process Design	420	10/74-10/75	270	1/74-8/74
- Owners Interests	380	10/75-7/77	390	8/74-9/75
- Startup and Test	550	7/77-4/79	500	9/75-4/77
	1,350		1,160	
• Contractors Design and Erection	1,200	10/75-7/77	1,000	8/74-9/75
	1,120			

(1) Cost of Engineering work only - costs of experimentation, pilot plant work, construction and demonstration plant operation are excluded.

(2) M = Thousands.

SECTION 81

REFERENCES

1. Study of Chemically Active Fluid Bed Gasifier for Reduction of Sulphur Oxide Emissions. Final Report, CAP Contract CPA 70-46, Esso Research Centre, Abingdon, Berkshire, June, 1972.

2. Curran, G.P., Fink, C.E. and E. Gorin, Phase II Bench-Scale Research on CSG Process, R&D Report No. 16. Report to Office of Coal Research, Contract No. 14-01-0001-415, Consolidation Coal Co. July 1st, 1969.

SECTION XII

INVENTIONS

1. UK 50614/72 — Moss, Craig, Taylor and Tisdall

   Preventing agglomeration during regeneration of sulphides by passing stone from the gasifier into a region of the regenerator which is separated by a layer of fluidised stone from the regenerator distributor.

2. UK 34739/72 — G. Moss

   Production of a highly sulphated lime from CaSO4 regenerator off-gas, to avoid the need for reduction of SO_2 to sulphur or production of sulphuric acid.

3. UK 23513/72 — Moss and Taylor

   Reduction of attrition in fluidised beds by a two stage nozzle, the first stage being a high pressure drop orifice, and the second stage providing dissipation of kinetic energy and a non-attriting gas velocity into the fluid bed.

SECTION XIII

GLOSSARY

TERMINOLOGY

Sulphur Differential	=	The difference in total sulphur level on the fluid bed between start and finish of a batch gasification run, or between the inlet and outlet streams from gasifier to regenerator in a continuous unit.
Superficial Velocity	=	The velocity of the fluidising gases (air plus flue gas recycle, but excluding gas and vapour formed from the fuel) in the empty gasifier or regenerator bed, at the temperature of the bed.
Fluidised Bed Depth	=	(Pressure drop from above distribution to gas space above the bed) ÷ Bed density
Lime Replacement	=	Fresh limestone added to the gasifier, expressed as weights of CaO in the limestone added over a given period per unit weight of sulphur in the fuel gasified during the same period. Alternatively expressed as a ratio or moles CaO added per mole S in the fuel gasified.
Sulphur Removal Efficiency(SRE)	=	$\left\{1 - \frac{SO_2 \text{ observed in flue gas}}{SO_2 \text{ if none absorbed}}\right\}$ x100%
Calcination	=	Removal of CO_2 from limestone by heating above approximately 750 deg. C.
Adiabatic Gasification	=	Operation at low air to fuel ratios in the gasifier such that heat released by partial oxidation of the fuel just serves to maintain the gasifier temperature at the required level.(Air supply about 1/3 of that needed to fully combust the fuel).
Combustion	=	Operation at high excess air levels during combustion in the gasifier such that the gasifier temperature just remains at the required level (Air supply about 400% of that needed to fully combust the fuel).

Megawatt (MW)	-	Used in this report only for electrical power generation rate. Thermal energy rates are all expressed in Btu/hr.

SYMBOLS USED IN TEXT

A	Bed age in hours (batch tests)
D	Bed Depth (inches)
d	Particle size, microns
d_{av}	Surface area mean particle size, microns
L	Loss rate from bed, g/min
S	Bed Sulphur Content (total), weight %
T	Bed Temperature, deg. C.
W	Weight of fraction in sieve analysis
μ	Micron (10^{-6} metre.)

SECTION XIV

APPENDICES

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APPENDIX A

Data From British Unit Studies

[illegible]

[illegible]

[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
[illegible]	1	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	[illegible]	[illegible]
[illegible]	2	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	3	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	4	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	5	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	6	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	–	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	7	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	–	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	8	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	9	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	10	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	11	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	12	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	13	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	14	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	15	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	16	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	17	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	18	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	19	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	20	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	21	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	22	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	23	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	24	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	25	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	26	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	27	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	28	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	29	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	30	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	31	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	32	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	33	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]
[illegible]	34	–	–	–	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	–	–	[illegible]	[illegible]

Table [illegible]

[illegible]

Time	[illegible]				[illegible]				[illegible]			
Time (hr)	Bed Depth (inches)	Bed Density [illegible]	Cyclone Dust [illegible]	Bed Temp. [illegible]	Bed Depth (inches)	Bed Density [illegible]	Cyclone Dust [illegible]	Bed Temp. [illegible]	Bed Depth (inches)	Bed Density [illegible]	Cyclone Dust [illegible]	Bed Temp. [illegible]
[illegible]	[illegible]	[illegible]	—	[illegible]	[illegible]	[illegible]	—	[illegible]	[illegible]	[illegible]	—	[illegible]
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
[illegible]	—	—	—	—	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
[illegible]	—	—	—	—	—	—	—	—	[illegible]	[illegible]	[illegible]	[illegible]
[illegible]	—	—	—	—	—	—	—	—	[illegible]	[illegible]	[illegible]	[illegible]
[illegible]	—	—	—	—	—	—	—	—	[illegible]	[illegible]	[illegible]	[illegible]
[illegible]	—	—	—	—	—	—	—	—	[illegible]	[illegible]	[illegible]	[illegible]

[illegible] Fines Return System

[illegible] gasifier [illegible] and [illegible] a system for externally treating the main gasifier [illegible] gas and returning the [illegible] recovered [illegible] the gasifier. [illegible]

[illegible] fines collection [illegible] is a typical [illegible] which can be isolated from its [illegible] by a butterfly valve A. A line leads from the bottom of each [illegible] cyclone to a common fines receiver mounted on the gasifier. A [illegible] injects pulses of N_2 into this [illegible] line at a 'Y' which is just beneath cyclone [illegible]. Another [illegible] gas enters each cyclone through valve [illegible].

The [illegible] line is isolated from the fines receiver by a ball valve B. The fines receiver projects into the side of an [illegible] vessel [illegible] and fitted with a plug [illegible] near its top. [illegible] the top of the [illegible], together with gas from the top of the fines receiver, [illegible] in which fines are [illegible]. The [illegible] from the bottom of the [illegible] through a water [illegible] heat exchanger [illegible] from which it falls into a sump [illegible] which is [illegible] to air inducer [illegible] the gasifier. A valve in the [illegible] line just above the air pickup is [illegible] by a differential pressure switch to prevent flow reversal in the [illegible] and the [illegible].

The operating sequence [illegible] valves A, [illegible], and B, is regulated from a control cabinet in the [illegible] room. There is a separate control panel and set of valves for each of the [illegible] gasifier [illegible].

The [illegible] control [illegible] elements performing all [illegible] and time delay functions. The sequence of operations of the [illegible] is as follows:

1. [illegible] A is open, valves [illegible] and B are closed. The vessel is filling with solids from the cyclone. A small [illegible] of N_2 into the body of valve B purges the cyclone drain leg of gasifier product gas.

2. When a timer resumes its mid-[illegible], a signal closes valve A. After a short delay, [illegible] A is [illegible], valve B opens to admit N_2 to the vessel to aerate and pressurize the solids in it. When the pressure reaches a suitable level, say 5 psi, valve B begins opening and closing in rapid sequence to inject pulses of N_2 into the transfer line. Simultaneously, valve B opens to deliver material to the fines receiver. Material flows from the vessel in broken into short slugs by the action of the N_2 pulses. These slugs travel with reduced pressure drop and with less gas consumption than would be required for conventional pneumatic transport.

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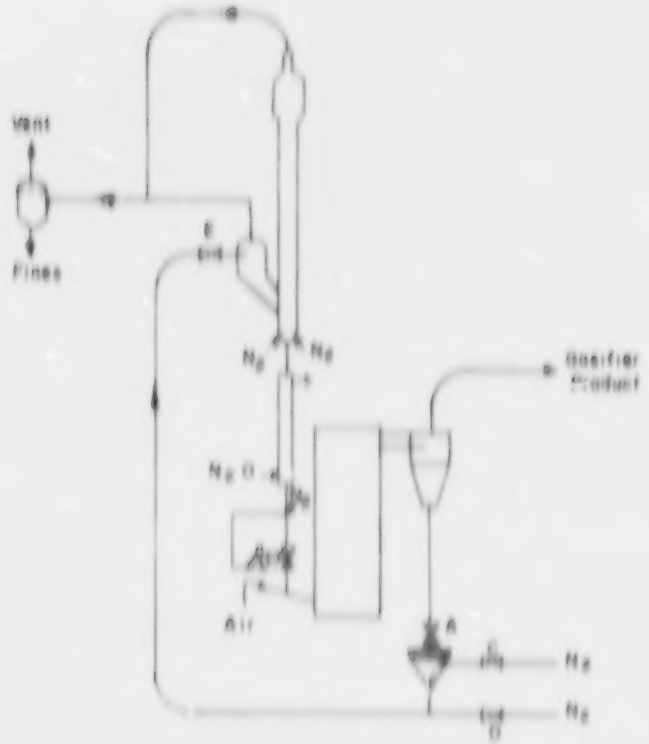
when the vessel is empty, the pressure within falls rapidly because there is no longer a resistance imposed by the solids. The pressure sensor detects this fall and closes valves C, D and E. Valve A then opens and a new filling period begins. We expect that fill times will be of the order of 15 minutes and emptying times about 1 minute or less. An interlock between the two parallel systems prevents both operating at the same time.

Solids drop from the fines receiver into the elutriator vessel where they are contacted with nitrogen at about 3 ft./sec. Fine solids are carried overhead by this gas while coarser material falls into the downcomer which also is fluidised by N_2. This downcomer is water cooled to reduce solids temperature enough to prevent rapid reaction with air. Fluidising N_2 for the downcomer is injected about one foot above the bottom outlet. Nitrogen from this injector passes both upward to fluidise the downcomer and downward to seal against backflow of air. A restriction at the bottom of the downcomer limits the flow rate of solids to prevent choking the air pickup. If the height of the fluid bed in the downcomer falls too low to provide an adequate seal against air backflow, the pressure detector senses the reduction and closes the bottom valve. The valve opens again when the pressure difference is restored.

A detailed description of the operation of the fluidic controller is given after Fig. B-1.



F.G.B.1 CYCLONE DRAIN SYSTEM

Description of Pneumatic Circuit

There are two parallel cyclone drain systems which empty to a common disposal. Each drain and transfer system is controlled by its own pneumatic control panel. The two systems are interlocked to prevent simultaneous transfer by both. The controllers are based on "Fluidyne" pneumatic elements made by Ideal Pneumatic Ltd., Telford, Shropshire. A diagram of the pneumatic circuit appears in Figure [illegible]. Table 1 compares the numbering system used in the circuit diagram with numbers on the physical panel.

Most of the logic elements in the circuit are 3 port valves with either pilot inlets at each end or a spring at one end and pilot pressure at the other.

The symbols used in this memo for these elements are as follows:

- If an element is energised left, it means that the left pressure is higher than right. In this case, the element lines are as follows:

- If an element is energised right, the opposite applies:

The numbers refer to the manner in which the element ports are numbered.

The symbol used at point 2 on the above example is a restrictive orifice. Point 3 is open to atmosphere with no restriction. Point 5 is plugged.

Element No. 1 in the diagram is a four port valve, air operated, spring loaded. With no air signal, it takes the configuration:

When air energised, it reverses to:

Elements [illegible] and [illegible] are two port valves, spring loaded. They are normally closed, that is, no flow passes through them unless pressure is applied to their pilot sides.

The oscillator consists of four elements and a regulator which causes pulses of air to be applied to the pilot of element 9 when a signal is applied to the oscillator. Details of this circuit are not illustrated here.

The figure illustrates the configuration of the valves when operating air at [illegible] psig has been applied to all points marked " [illegible] ", but the switch is in the "off" position.

At this point there is no pressure on either pilot side of element [illegible]. The position shown was maintained from the previous operation by a "detent" feature which prevents movement of the element when pressure is off. The element is normally in this position except for the brief period during which tank pressure is building before the outlet valve has opened.

The sequence of operations is listed in Table [illegible] and described briefly below.

Pressure is supplied up to the switch from element [illegible] through element [illegible], but is blocked to the pilot of [illegible] and inlet valve [illegible] by the switch. Thus the inlet is closed while the switch is off.

When switch is turned on, pressure is applied to [illegible] to open the inlet valve and to reverse [illegible] to pressurized the [illegible] delay accumulator and start [illegible] accumulator exhausting through an orifice. The exhausting of [illegible] supplies the main time delay for solids to drain from the cyclone to the transfer tank. This time is expected to be of the order of 10-15 minutes.

Because of the pressure applied through [illegible] to the pilot of [illegible], it reverses. At this point there is still no pressure on either side of [illegible], but it remains held in position by a "detent" spring. The pilot of [illegible] is pressurized by the air pressure to valve A, and it reverses. Element [illegible] remains held by "detent" with no pressure on either side.

When the pressure in [illegible] accumulator drops below that supplied by "delayed" regulator, element [illegible] reverses. This starts [illegible] exhausting and pressurizes the pilot of [illegible] which reverses. Pressure is removed from [illegible] pilot which reverses to close the inlet valve.

Pressure is released from [illegible] pilot which reverses in readiness to supply pressure to the left pilot of [illegible].

When [illegible] pressure falls below that of [illegible] regulator, [illegible] reverses. This delay was provided to give inlet valve about [illegible] seconds to close fully. Pressure now is supplied through [illegible] to the left pilot of [illegible] which reverses. Pressure from [illegible] then goes through [illegible] reverse [illegible]. Pressure from [illegible] also reverses [illegible] to cut "x" pressure on the pilot of [illegible]. The reversal of [illegible] applies pressure from [illegible] to the pilot of [illegible] which opens [illegible] admitting nitrogen to the transfer tank aerator. Element [illegible] supplies this signal to operate its aerator as long as the second system is not operating the other tank aerator. If the other system is operating, the signal to open valve [illegible] is delayed until transfer from the other system is completed. Likewise, a signal from this system prevents the second from starting if the first is operating. The purpose of this interlock is to prevent simultaneous high nitrogen demand from both transfer units.

While the transfer tank is building pressure, the outlet valve remains closed until tank pressure exceeds "x" pressure, say 5 psig. When this happens, element 7 reverses to apply a signal to the oscillator, to element 8, and to the right side of 12. Element 7 removes the signal from right of 3 and from 10. The oscillator starts pulsing N_2 through the knife via 5. Element 8 reverses to open the outlet valve. Solids should now start transferring from the transfer tank to the elutriator. The element 10 reverses to change the signal to 11 which reverses to apply "x" pressure to the right pilot of 7.

When transfer of solids is completed, tank pressure falls quickly. When it drops below pressure "x", element 7 reverses to restore conditions to the starting point. Signal is removed from the oscillator, from 8, and from the right side of 12. The oscillator stops pulsing the knife through 5 and element 8 reverses to close the outlet valve. Pressure is applied through 7 to 3 which reverses to stop the N_2 flow to the aerator, and signal is applied to 10 and to 1 which reverses to open the inlet valve. Reversal of 10 starts the long time delay exhausting of R_{22} and repressurizes R_{21} to reverse element 11. The transfer tank is now again receiving solids from the cyclone.

Table 4.1

Composition of Element Numbering System

Number on Fig.1	Number on Panel	Catalogue No
1	17	PIV 50/168/B
2	16	PIV 50/168/B
3	- (on front)	
4	9	PIV 50/138/B
5	11	PIV 50/158/B
6	- (in pit)	
7	13	PIV 50/168/B
8	15	PIV 50/138/B
9	- (in pit)	
10	18	PIV 50/158/B
11	164	PIV 50/138/B
12	7	PIV 50/168/B
13	12	PIV 50/177/B
14	behind panel	PIV 50/141/B

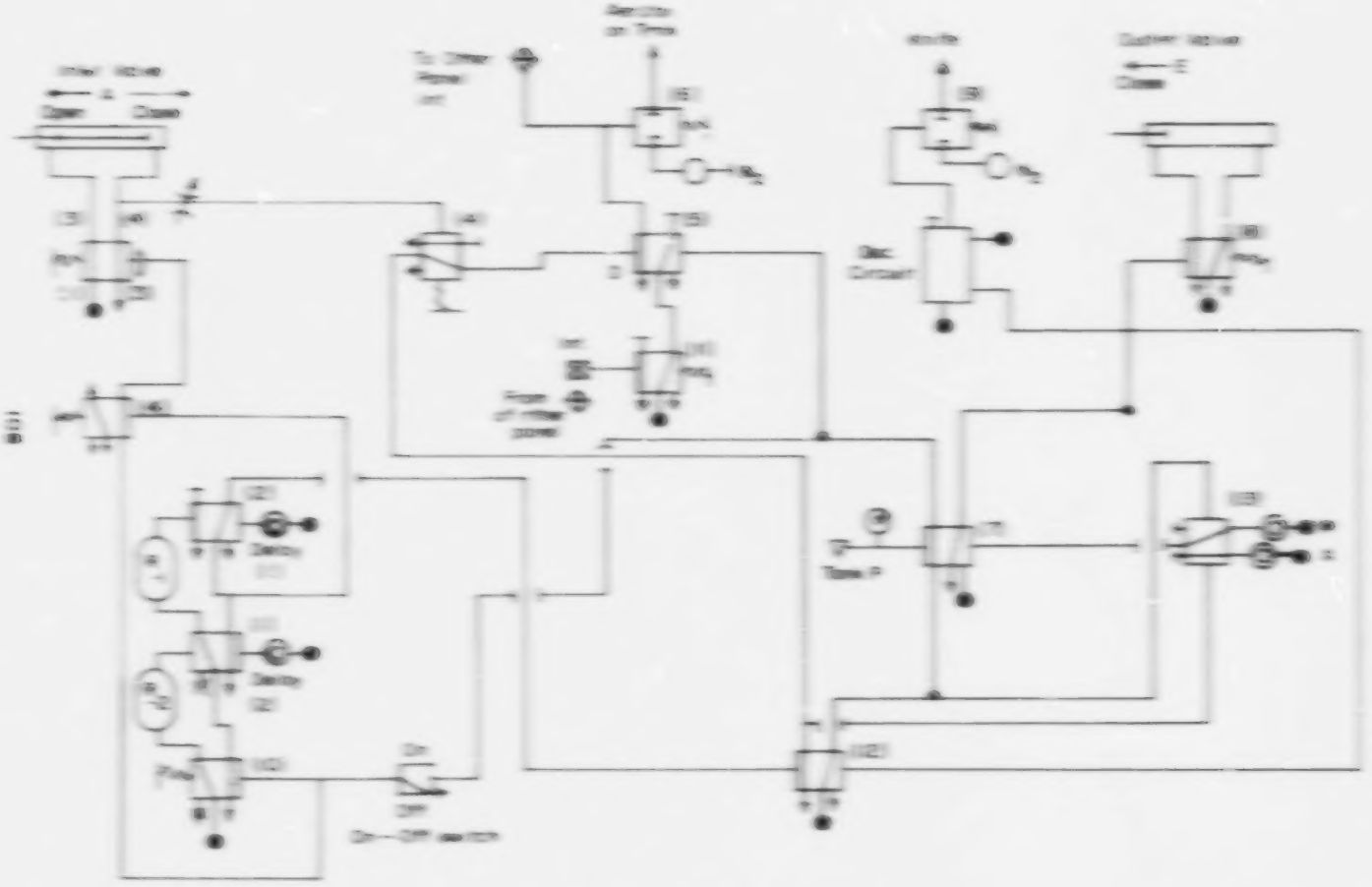


FIG B-2 CIRCUIT OF CYCLONE FINES PUMP CONTROLLER

APPENDIX 1

Examination of the CAFB Gasifier and Auxiliary Equipment

Gasifier and Regenerator Shell

Gasifier Concrete

The walls of the gasifier were generally blackened with carbon and in the upper regions around the junction between flue lid and the walls the carbon was up to ¼" thick. There were patches immediately above the cyclone inlet ducts where the carbon had burnt away suggesting that an air leak has occurred after the shut down. In the lower areas, the wall was glazed with a hard thin tenacious carbon deposit.

The cracks in the concrete hot faces which have been present from the first firing of the unit showed their customary fine black deposit of carbon about 1" wide in the upper areas of the wall. The lower areas were clear because of the splashing action of the bed material. There was no sign of any deterioration in the concrete from this test run.

There were areas at the junctions between the distributor and the walls where a deposit of fine material had agglomerated together to form a small coving between the distributor and the wall, most likely caused by an area of poor fluidisation due to a blocked or partially blocked distributor nozzle.

Gasifier Penetrations

The thermocouples, fuel injectors and pressure tappings were in good order throughout although the left hand fuel injector had been replaced during the run.

The thermocouple in the lid had a considerable growth of carbon around its protruding end whilst those in the bed area were generally clean, apart from the one at the lowest point in the bed close to the fines return pipe. There was some agglomeration of lime and carbon bridging between this thermocouple, the distributor and the gasifier wall. This condition probably arose from poor local fluidisation due to an obstructed nozzle in the distributor together with the introduction of the fines from the return system into this poorly fluidised zone. This thermocouple did show a consistently low reading during the latter part of the run which was probably caused by this local static material.

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The fuel injectors were layered with carbon on their protruding sections and the injector at the right hand side had a hollow cap of carbon and lime enclosing its end.

The two air injection tubes which pass through the lid to direct air into the cyclone inlets were both heavily scaled and burnt away at their protruding tips.

The stainless steel tubes for the stone feed and fines return were both in good order.

Cyclones

The cyclone bodies and their inlet sections which had both been lined with type 310 stainless steel to improve their surface finish and hence performance were heavily obstructed with a mixture of carbon and lime.

The inlet ducts are illustrated in Figure 6.1 and Figure 6.2 which show the black deposits around the inlets which gradually become lighter further into the cyclone.

The white area on the gasifier wall immediately above the cyclone inlets can only be explained by some areas of carbon burning out after the unit was shut down possibly due to some local movements which may have broken any seals existing between the underside of the lid and the gasifier wall.

Figure 6.1 also shows the white irregular deposits on the cyclone outlet tube which can be seen hanging down inside the cyclone body.

Figure 6.3 shows a view into the right hand cyclone after removal of the outlet tube shown in figure 6.4. The cyclone upper body was obstructed around its total circumference leaving the hole in the centre formed by the outlet tube. The material was laid down in a very irregular manner consisting of layers and folds of fine hard material with a tortuous gas path amongst this deposit. The deposit becomes less pronounced towards the bottom of the cyclone compared to the upper section, but still filled a considerable volume of the lower section. The stainless steel liner was badly scaled and distorted, in some areas it was completely burnt away due to the high surface temperatures when carbon was burnt off at the various burn out operations during the run. In the upper sections of the cyclone the liner was soft and easily broken into flakes but nearer the bottom of the cone the lining was generally stronger and some quite large pieces of steel were removed intact. The steel lining tube which sealed off the cyclone drain passage to the gasifier to recirculator transfer line was heavily scaled with a small area burnt away at the top retaining collar.

The deposit at the intersection of the rectangular inlet with the cylindrical cyclone body was layered with white fine material separated by thinner blacker layers, suggesting that the thicker white layers are laid down during some of the combusting periods. This is confirmed by the batch unit tests which showed that combusting conditions give rise to the most severe material deposition.

The silicon carbide outlet tube of the right hand cyclone shown in figure 6.5 was removed with very little material adhering to the outer surface. Inside the tube there was an overall thin layer of carbon and lime about 1/16" thick deposited around the bore and in some areas near the bottom of the tube there were thicker irregular deposits up to 5/8" thick. These thinner layers could be removed fairly easily but the thinner layers were very firmly bonded to the tube.

The left hand cyclone was also severely blocked in the inlet section and in the body of the cyclone although it was not as severe as the right hand cyclone (figure 6.6). The cyclone outlet tube was coated on its outside with a thick black [illegible] deposit (figure 6.7) but the deposits did not completely bridge across the gas passage between the outlet tube and the cyclone wall.

The bore of the tube was coated with a layer of carbon and lime about 1/16" thick deposited fairly evenly over the surface of the tube.

The stainless steel cyclone liner was severly coated and locally distorted or burnt away in many locations near the top of the cyclone. Towards the lower end the lining was almost intact although still heavily coated.

The stainless steel tube sealing off the drain to the regenerator to gasifier transfer line was heavily coated and around the upper outside surface there were crystalline deposits of carbon which could have come up the transfer line from the gasifier bed. The upper retaining collar around the tube had burnt completely away.

Gasifier Distributor

The distributor was generally in good condition apart from some obstruction in the holes in the nozzles. The obstructions arose from two sources : one was from lime particles which entered from the gasifier bed and the other deposits have come from fine rust scale carried through from the flue gas recycle line which was carrying a saturated gas from the water scrubber.

The distributor nozzle design has three staggered holes in series with each other, the first smaller one to provide the pressure drop needed in a fluid bed distributor and the third larger one provides a low exit velocity to minimize damage to the stone in the bed. The middle hole forms a plenum between the inlet and outlet holes.

Generally, the outer holes were obstructed more with lime particles and the inner holes with deposits of rusty coloured material. There were considerable problems encountered in removing the distributor because it had become welded with limestone and heavy fuel oil when one of the fuel injectors failed during the run. Some of the mechanical force used to remove the distributor may have dislodged some of the material fouled in the nozzles but examination did show that 52 of the 100 outlets were obstructed completely; 10 of the inlet holes were completely blocked, 100 partly obstructed and the remaining 65 holes were clear.

The stainless steel material used for the nozzles was in excellent condition and showed no sign of deterioration.

Gasifier Lid

The lid was heavily coated with carbon on its lower face and the protruding thermocouple had a considerable deposit of carbon around it. The refractory concrete was in good condition but the vermiculite and calcium silicate insulating slab was cracked in a number of places.

Gasifier Bed Material

The gasifier bed was slumped without sulphation at shut down and figure 6.7 shows the typical bed material after half the material had been removed. There are two interesting features in this picture, one of them being the random area of completely white material which proved to be an area of fine particles free of carbon obviously formed after the bed was slumped which must then have had some oxygen in-leakage to burn off the carbon. This area of carbon free material existed in about the middle third of the bed depth. Another interesting and typical feature is the blackened area at the right hand side which corresponds to No.1 fuel injector location. After further bed removal this fuel injector was found to be encased with a hollow sphere about 5" diameter of carbon and lime particles which would have restricted the throw of the injector.

The material was generally free from any large agglomerates apart from a few lumps of carbon and fine lime particles up to 1" across. There were some areas where there had been stable zones in the fluid bed around the periphery of the distributor particularly near the fines return pipe from the cyclone shown in the right hand side of fig. 6.6.

Regenerator

The regenerator material was free from agglomerates and the bore of the regenerator was generally clean apart from a few small deposits at the top outlet section and around the joint between the distributor and the wall. The refractory did not show any deterioration.

The distributor was in good condition with the nozzles clear and undamaged. The refractory lip around the distributor had cracked away during the run and the repair that was made had withstood the remainder of the run without any deterioration.

The transfer passages to and from the regenerator were free of any agglomerates and the refractory concrete around the transfer sections was in good condition.

Ductwork and Burner

Gasifier Outlet and Burner

The bifurcated duct had a uniform deposit of carbon on its inner surface about 1/16" thick and at the junction between the two ducts there were thicker deposits where the two gas streams converged. The thermocouple in this area was heavily coated with a carbon and lime layer about ½" thick.

The premix section upstream of the burner had a 1/16" carbon layer on the stainless steel section and the 1/8" dia. stainless steel thermocouple in the burner throat was burnt away.

The main burner was in good shape with a layer of carbon about 1/16" thick deposited on the internal gas ductwork. The outlet ring of the burner had a local build up mainly of lime on its outer face shown in fig 6.3 which was deposited from the turbulence in the gas streams arising from the stainless steel baffle plates located to shield the pilot flame from the main combustion air.

The pilot burner which has been very successfully modified to provide a forced gas and air premix system prior to the flame retention nozzle was in good condition apart from some local lime deposits around the nozzle and which protruded into the burner quarl in the boiler.

The stainless steel deflector plates in the boiler burner quarl were heavily deposited with lime.

Regenerator Outlet

The outlet pipe from the regenerator top gas space was coated with long thin light purple deposits of material laid along in the direction of flow and projecting out from the wall of the duct (figure 6.10). In one area there was one projection which extended out almost 4" from the wall but generally they were ½" or less and only about 1/8" thick at the narrowest tip. They were held fairly firmly to the wall and appeared to have passed through a molten phase having a fairly smooth outer face. Further downstream the pipe was coated with a much more uniform deposit about 1/8" thick which was composed of fine particles rather than the liquid type of deposition in the hotter upstream section of the pipe (Figure 6.11).

Downstream of the dust extraction cyclone the pipe was clean apart from a very thin white fine deposit.

Boiler and Stack

Boiler

The back end of the boiler had been cleaned periodically during the test run and figure 0-12 shows the condition after shut down. There was some coarser material laying in the bottom of the boiler fire tube with some finer material deposited around the lower sides of the fire tube mainly on the left hand side when viewed from the rear, indicating that there was some swirl in the flame at the burner.

The entries to the first pass of fire tubes were deposited with rings of fine material although none of the tubes was totally blocked. Some of the deposits were smooth and rounded whilst others were spiky, stick shards in figure 0-13.

The return tube passes were generally clear apart from a few tubes which had some deposits at their ends. Some material had been deposited out of the gas stream and collected in the boiler space at the end of this second tube pass.

The refractory on the boiler rear door showed some signs of pitting but there has been a gradual deterioration not just associated with this particular run.

Boiler Probe

The boiler probe acquired some local light brown lumpy deposits which tended to be more concentrated around the "snout" of tube figure 0-14, in front of the entrance to the first tube pass. There was no indication of any deterioration of the tube which was cooled during the run to approximately 200°C.

Stack and Cyclone

There was a quantity of fine at the base of chimney which was otherwise clean and the external cyclone and knock out pot were also clear of any obstruction. Both these vessels were continually rapped during the run and this prevented any significant build up of material.

Flue Gas Scrubber and Recycle Line

Scrubber

There were considerable operational problems with the scrubber initially due to the wet fine slurry which was discharged and the deposits which built up at the scrubber entry. The problems were eased by running the scrubber with some recycle so reducing the dust burden concentration and increasing the velocity of the gas and hence efficiency of operation.

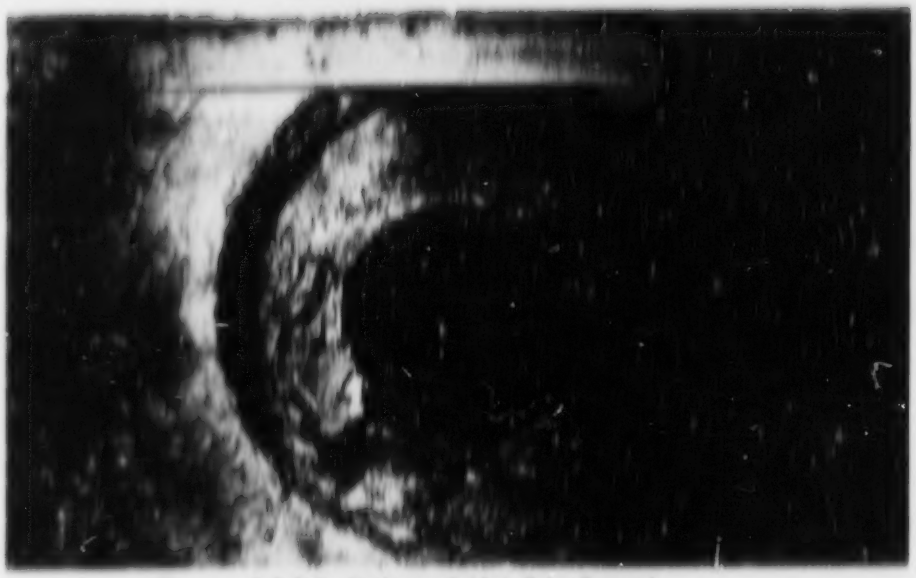
Examination afterwards showed that the knock out vessel had some hard fine deposits on its wall opposite the gas entry when the material would first impinge on the wall. The scrubber and its entry pipe was clear.

Recycle Line

The scrubber has cleaned up the solids content of the gas very well but introduced some problems due to the cold saturated gas leaving the scrubber which contained enough very fine particles to seize up the cycle line control butterfly valve and the control valve to the burner air valve which was in a static leg and probably contained a considerable quantity of condensation.

Figure C-1 R.H. Cyclone Inlet

Figure C-2 L.H. Cyclone Inlet



C-3 R.H. Upstream, Outlet Tube Removed

Figure C-3 R.H. Upstream Outlet Tube

U-3 L.H. Cyclone, Outlet Tube Removed

Figure U-3 L.H. Cyclone Outlet Tube

Figure C-7 Oscillator Box Half Removed

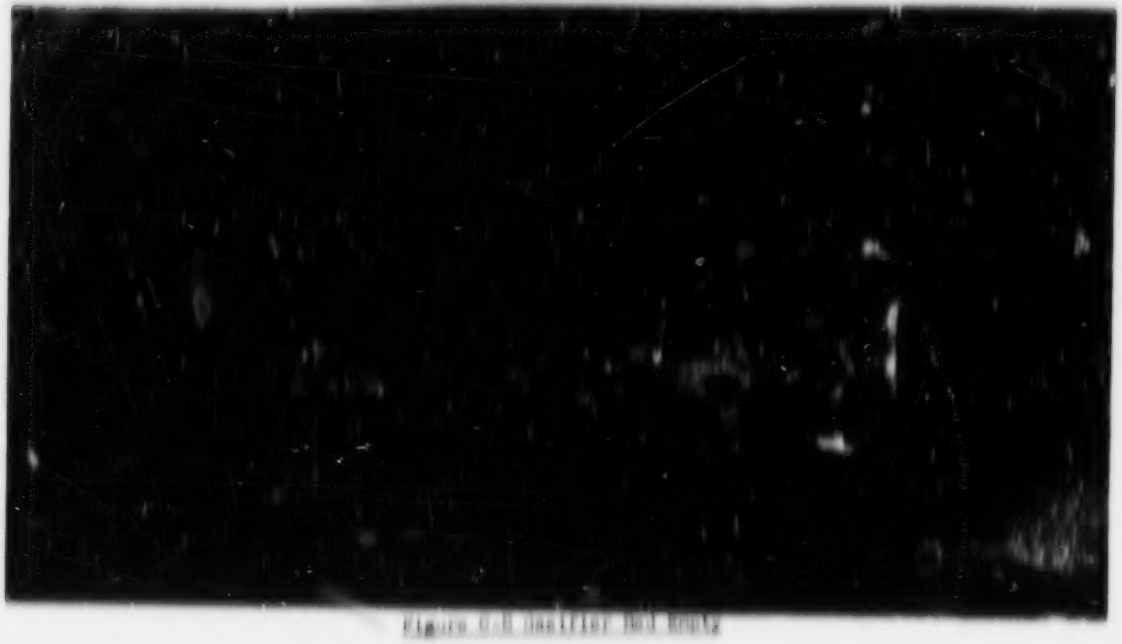


Figure C-8 Oscillator Box Empty

Figure C-9 Main Burner Outlet Ring

Figure C-10 Regenerator Outlet Pipe, Upstream End

Figure C-11 Regenerator Outlet Pt/Y, Downstream End

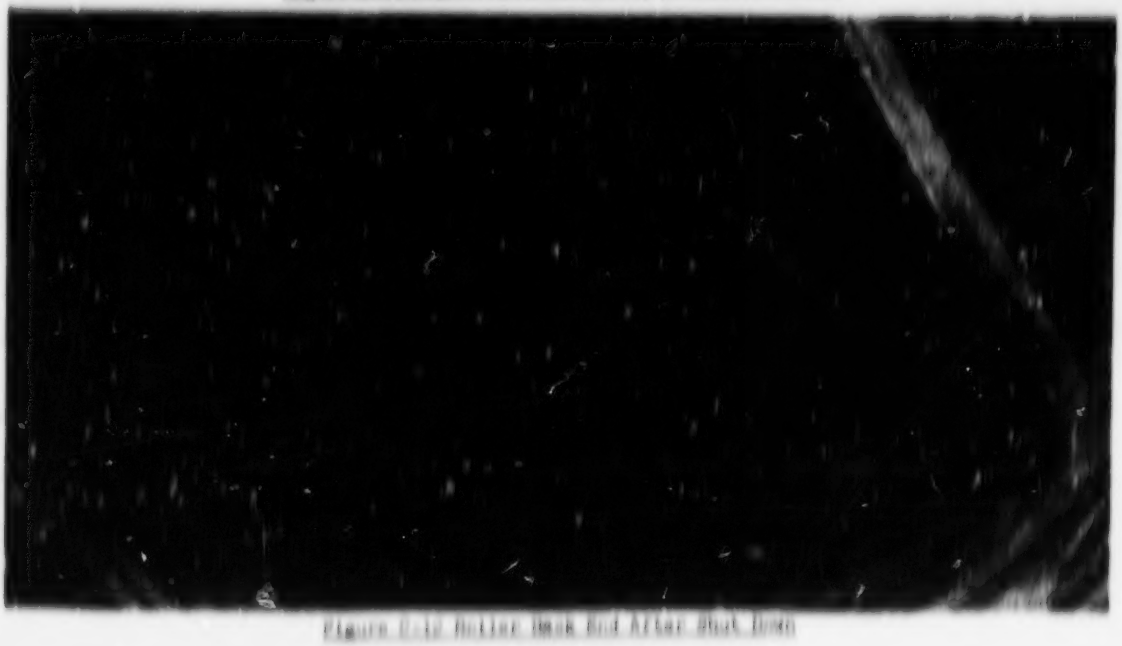


Figure C-12 Boiler Hook End After Shut Down

Figure C-13 Fire Tube First Pass Inlets

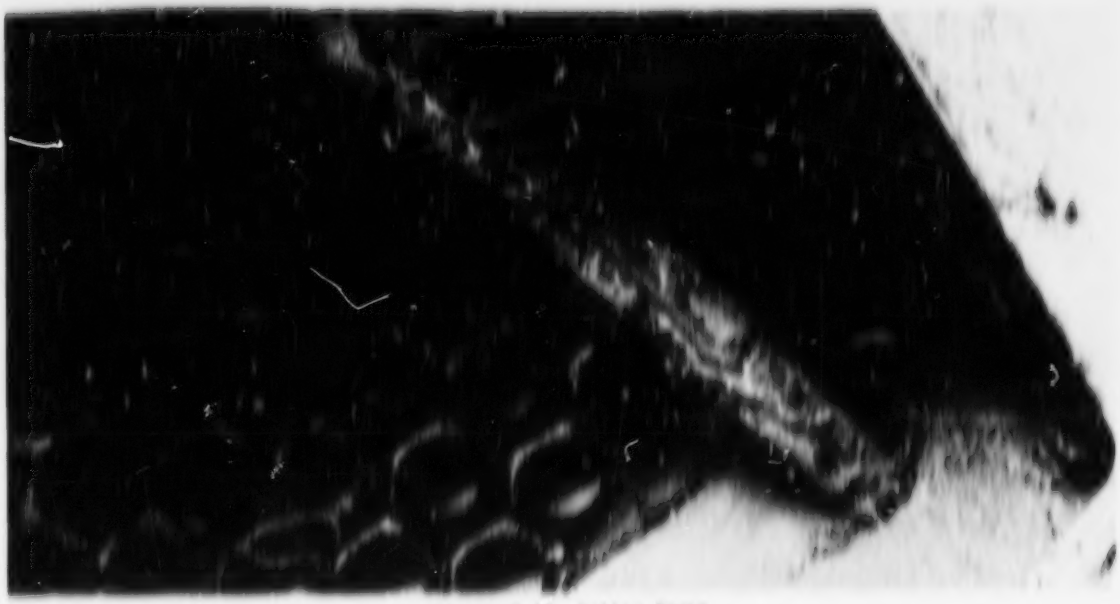


Figure C-14 Boiler Front

APPENDIX D

Data for Run 3

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Table D.1

Run 1 Denbighshire Blue

CARB GAS COMPOSITIONS

DAY/HOUR	FLUE GAS CO2 %	FLUE GAS CO* %	FLUE GAS FPM	REGENERATOR GAS CO2 %	REGENERATOR GAS CO* %	REGENERATOR GAS SO2 %	CASSIFIER INLET GAS VOL % CO2 ANAL	CASSIFIER INLET GAS VOL % CO2 CALC	CASSIFIER INLET GAS VOL % CO ANAL	CASSIFIER INLET GAS VOL % CO CALC
0.0130	0.0	18.6	989	0.1	0.5	0.7	16.5	16.3	0.7	0.7
0.0930	0.2	18.6	987	0.1	0.6	0.8	16.5	16.3	0.6	0.7
0.0930	0.1	17.2	987	0.1	0.6	0.9	15.0	16.9	0.6	0.5
0.0630	0.7	13.8	958	0.1	0.5	0.9	15.0	15.1	0.3	0.6
0.0530	0.0	13.5	58	0.1	0.9	0.9	15.0	15.0	0.7	0.3
0.0630	0.5	13.7	97	0.1	0.5	0.3	15.0	15.6	0.1	0.9
0.0730	0.5	13.5	977	0.1	0.7	0.6	15.0	15.0	0.0	0.1
0.0930	0.1	18.6	758	0.1	0.6	0.5	15.5	16.0	0.8	0.0
0.1030	0.0	18.6	765	0.1	0.6	0.6	16.0	16.9	0.6	0.9
0.1130	0.0	15.6	585	0.1	1.7	0.7	16.0	11.5	0.6	7.0
0.1930	0.5	15.3	518	0.1 0.0	0.6	0.6	16.0	13.6	0.5	5.0
0.1130	0.0	18.8	563	0.1	0.6	0.1	16.5	16.5	0.3	5.0
0.1830	0.0	15.0	693	0.1	1.3	0.0	16.0	16.0	0.5	5.0
0.1530	0.0	18.7	673	0.1	1.3	0.6	16.5	16.6	0.3	5.1
0.1630	0.1	18.6	618	0.1	1.1	0.6	16.5	16.6	0.1	5.0
0.1730	0.0	18.6	765	0.1	1.0	0.3	16.5	16.6	0.1	0.0
0.1830	0.1	18.7	763	0.1	0.8	0.0	16.0	16.6	0.0	5.0
0.1930	0.0	18.7	936	0.1	0.8	0.0	16.0	16.6	0.0	5.0
0.2030	0.5	18.1	936	0.1	0.6	0.1	15.0	15.6	0.0	0.3
0.2130	0.0	18.7	936	0.1	1.5	0.0	16.0	16.6	0.3	0.1
0.2230	0.1	18.6	818	0.1	1.3	0.0	16.5	16.9	0.0	0.7
0.2330	0.0	18.6	818	0.1	1.0	0.6	15.0	15.3	0.0	0.6
0.2830	0.0	18.6	818	0.1	1.0	0.0	15.0	15.0	0.0	0.3
1.0130				0.1						
1.0230	0.0	18.6	818	0.1	1.0	0.6	15.0	15.6	0.3	0.0
1.0430	0.0	18.1	800	0.1	1.3	0.3	15.0	15.3	0.0	0.1
1.0630	0.1	18.6	800	0.1	1.3	0.1	15.5	15.6	0.0	0.0
1.0730	0.1	18.1	809	0.0 0.5	1.0	0.6	15.5	15.1	0.5	0.0
1.0830	0.1	18.7	981	0.0 0.5	1.1	0.6	15.5	15.9	0.5	0.0
1.0730	0.1	18.6	800	0.1	1.1	0.1	15.5	15.9	0.5	0.1
1.0930	0.0	18.7	918	0.1	1.3	0.1	16.0	15.9	0.5	0.0
1.0930	0.0	18.6	799	0.1	1.3	0.1	16.0	16.0	0.6	0.5
1.1030	0.1	18.6	990	0.1	1.1	0.0	15.5	15.1	0.0	0.0
1.1130	0.0	18.6	985	0.1	0.6	0.1	15.0	16.9	0.5	0.1
1.1530	0.0	15.6	956	0.1	0.6	0.3	15.0	15.1	0.3	0.5
1.1130	0.0	18.7	918	0.1 0.0	0.6	0.0	16.5	16.6	0.3	0.6
1.1630	0.5	17.8	838	0.1	1.7	0.6	16.5	16.7	0.1	0.7
1.1530	0.5	18.6	880	0.1	0.6	0.0	15.5	16.3	0.1	0.0
1.1630	0.0	18.6	963	0.1	0.6	0.0	15.5	16.3	0.0	0.0
1.1730	0.0	18.6	870	0.1	1.0	0.0	15.0	16.5	0.0	0.0

Table 7-1

Run 3 Denbighshire Runs

CAER GAS COMPOSITIONS

DAY:HOUR	FLUE GAS O2 %	FLUE GAS CO2 %	FLUE GAS SO2 PPM	REGENERATOR GAS O2 %	REGENERATOR GAS CO2 %	REGENERATOR GAS SO2 %	DASIFIED ISLET GAS VOL % O2 ANAL	DASIFIED ISLET GAS VOL % O2 CALC	DASIFIED ISLET GAS VOL % CO2 ANAL	DASIFIED ISLET GAS VOL % CO2 CALC
1:1530	0.8	13.5	470	0.1	1.6	0.1	15.5	15.1	1.0	0.6
1:1930	0.8	13.5	563	0.1	1.6	0.1	15.5	15.1	1.0	0.6
1:2030	0.9	14.4	419	0.1	0.9	1.1	15.5	14.9	0.9	0.6
1:2130	0.1	14.7	600	0.1	0.6	1.1	15.0	14.9	0.9	0.6
1:2230	1.7	14.7	581	0.1	1.6	0.3	15.0	14.7	0.9	0.7
1:2330	0.1		681	0.1	0.9	1.8	15.1	14.8	0.9	
2:0030	0.1		800	0.1		1.9	15.0	14.8	0.9	
2:0130	0.3	14.8	615	0.1	1.1	0.6	15.0	15.5	0.8	0.7
2:0230	0.6	14.8	610	0.1	1.0	5.0	15.0	15.6	0.7	0.0
2:0430	0.6	14.1	600	0.1	1.3	0.3	15.0	15.7	0.6	0.1
2:0630	0.3	14.8	600	0.1	1.9	0.1	15.0	15.6	0.6	0.1
20:2030	0.0	15.0	70	0.00	1.3	0.5	18.0	17.9	1.6	0.5
20:2130	1.6	15.6	50	0.00	1.3	1.6	17.5	17.1	0.9	1.0
20:2230	1.6	15.6	70	0.00	0.6	1.0	17.7	17.0	0.1	1.0
20:2330	0.0	15.3	65	0.00	0.0	1.0	18.0	17.3	1.0	1.0
21:0030	0.0	15.0	57	0.00	1.1	1.0	19.0	18.5	1.1	0.0
21:0130	0.0	15.0	10	0.00	1.5	1.9	19.0	18.5	1.0	0.0
21:0230	0.0	15.0	10	0.00	1.3	0.9	19.3	18.5	1.1	1.9
21:0130	1.0	15.0	65	0.00	0.0	1.0	18.7	18.1	1.0	0.0
21:0630	1.0	15.3	9	0.00	1.9	1.9	18.0	18.1	1.3	0.3
21:0530	0.0	15.3	9	0.00	0.6	0.6	18.0	18.1	1.3	0.3
21:0630	0.0	15.3	100	0.00	1.0	1.9	19.0	18.0	1.3	0.3
21:0730	0.3	15.0	109	0.30	0.0	1.0	19.0	18.1	1.3	0.3
21:0830	0.0	15.0	109	0.00	1.7	0.3	19.0	18.1	1.3	0.3
21:0930	0.0	15.0	109	0.00	1.9	1.9	19.0	18.0	1.3	0.3
21:1030	0.0	14.7	101	0.00	1.9	1.6	19.0	18.0	1.0	0.3
21:1130	1.5	15.3	58	0.10	0.5	0.1	19.0	18.0	1.3	0.6
21:1030	1.0	15.3	50	0.10	0.7	0.7	19.0	17.9	1.3	0.6
21:1330	1.0	14.6	136	0.10	1.3	0.3	19.0	18.1	1.0	0.0
21:1630	1.5	15.3	107	0.10	1.3	0.3	19.0	17.9	1.0	0.6
21:1530	1.0	15.3	81	0.10	1.7	1.6	19.0	17.9	1.0	0.6
21:1630	1.0	15.0	58	0.10	0.3	0.9	19.0	17.9	1.0	0.3
21:1730	1.3	14.7	165	0.10	0.3	1.9	19.5	18.0	1.0	0.3
21:1830	1.0	15.3	000	0.10	0.6	5.1	19.0	17.9	1.0	0.9
21:1930	1.0	15.3	000	0.10	0.9	5.0	00.5	18.0	0.0	0.6
21:2030	0.0	14.7	163	0.10	0.6	1.9	17.0	15.8	0.6	0.0
21:2130	0.0	14.7	510	0.00	1.5	1.6	15.0	14.0	1.3	0.7
21:2030	1.7	15.0	265	0.00	0.5	1.0	15.7	14.1	1.3	0.0
21:2130	1.1	15.0	150	0.00	0.1	1.6	16.0	15.5	0.9	0.3
22:0030	0.0	15.0	004	0.1	0.5	1.0	16.5	15.6	0.9	0.3

Table D-1

Run 1 Pembrokeshire Stone

CAPR GAS COMPOSITIONS

	FLUE GAS			REGENERATOR GAS			ORIFICE INLET GAS			
DAY:HOUR	O2 %	CO2 %	SO2 PPM	O2 %	CO2 %	SO2 PPM	VOL % O2 ANAL	VOL % O2 CALC	VOL % CO2 ANAL	VOL % CO2 CALC
09:01:30	3.1	15.0	180	0.50	2.6	8.1	16.7	15.6	0.0	0.3
09:02:30	0.0	18.3	81	0.30	2.6	1.8	18.0	15.6	0.0	0.0
09:03:30	0.0	15.1	300	0.30	5.3	0.3	16.8	15.6	0.0	0.3
09:04:30	0.1	15.0	185	0.00	0.5	1.6	16.8	15.6	0.0	0.3
09:05:30	1.0	15.0	185	0.00	0.6	0.3	16.6	15.6	0.0	0.3
09:06:30	2.6	15.1	197	0.00	1.1	0.6	16.9	15.7	0.1	0.3
09:07:30	0.5	18.7	136	0.00	0.9	1.6	17.0	15.0	0.6	0.0
09:08:30	0.5	18.7	189	0.1	5.0	0.6	16.5	15.6	1.0	0.3
09:09:30	0.5	18.7	136	0.10	1.6	1.6	17.5	15.7	0.0	0.0
09:10:30	0.0	15.0	183	0.10	1.7	0.9	16.5	15.6	0.0	0.0
09:11:30	0.0	15.0	181	0.1	0.6	1.6	16.5	15.6	0.0	0.0
09:12:30	0.5	18.8	181	0.1	5.0	1.9	16.5	15.7	0.6	0.1
09:13:30	0.0	18.8	309	0.10	3.1	0.9	16.5	15.6	0.1	0.1
09:14:30	0.5	18.7	304	0.10	0.3	0.3	16.5	15.7	0.7	0.0
09:15:30	0.5	18.7	081	0.10	3.3	0.3	16.5	15.7	0.6	0.0
09:16:30	0.3	18.7	190	0.10	3.3	1.7	16.5	15.6	0.6	0.0
09:17:30	0.5	18.8	310	0.10	3.6	0.3	16.5	15.7	0.6	0.1
09:18:30	0.0	18.5	307	0.10	3.3	0.9	16.5	15.6	0.7	0.1
09:19:30	0.0	18.7	110	0.30	1.0	1.6	16.1	15.0	3.1	5.6
09:20:30	0.5	18.7	156	0.00	0.0	0.0	16.0	15.0	3.1	5.7
09:21:30	0.5	18.7	110	0.10	0.0	3.9	16.0	15.0	3.1	5.0
09:22:30	0.5	18.7	0	0.00	1.1	1.1	15.0	15.0	3.1	5.1
09:23:30	0.5	15.0	110	0.50	0.0	0.5	15.0	15.0	0.0	0.0
09:24:30	0.3	18.7	110	0.50	0.0	1.0	16.2	15.0	3.1	5.1
09:25:30	0.0	18.7	110				16.2	15.0	3.9	5.1
09:00:30	2.6	15.0	11	0.10	1.5	0.3	16.0	15.0	0.0	0.0
09:01:30	0.6	18.6	11	0.10	1.7	1.6	16.5	15.0	0.9	0.3
09:02:30	0.3	18.8	10	0.10	1.0	1.9	16.5	15.1	0.5	0.5
09:03:30	0.5	18.8	15	0.1	0.5	1.2	16.5	15.0	0.6	0.5
09:04:30	0.0	15.0	15	0.1	0.5	1.6	16.5	15.0	0.6	0.3
09:05:30	0.0	15.0	0	0.1	1.5	1.1	16.5	15.0	0.3	0.6
09:06:30	0.3	15.0	0	0.1	0.0	1.6	16.5	15.0	0.5	0.6
09:07:30	0.0	18.9	0	0.1	1.3	0.6	16.7	15.0	0.6	0.7
09:08:30	0.3	15.0	0	0.1	1.3	0.3	17.0	15.0	0.6	0.0
09:09:30	0.3	15.0	0	0.1	1.3	0.9	17.0	15.0	0.6	0.0
09:10:30	0.0	18.7	0	0.10	1.6	0.9	17.0	15.1	0.0	0.0
09:11:30	1.0	18.7	0	0.00	0.5	1.9	16.5	15.5	0.6	0.5
09:12:30	0.1	18.7	0	0.00	0.0	1.6	17.0	15.1	0.6	0.5
09:13:30	0.1	18.7	0	0.10	1.1	0.3	16.1	15.0	0.0	0.4
09:14:30	0.1	18.7	55	0.1	0.6	1.9	16.0	14.9	0.0	0.3

Table D-1

Run 3 Hemisphere Brand

CAER GAS COMPOSITIONS

	FLUE GAS			REGENERATOR GAS			GASIFIER INLET GAS			
DAY:HOUR	O2 %	EOP %	SO2 PPM	O2 %	COP %	SO2 %	VOL % O2 ANAL	VOL % O2 CALC	VOL % EOP ANAL	VOL % EOP CALC
7a:1530	0.0	15.0	56	0.1	0.0	3.9	16.0	14.9	0.0	0.0
7a:1630	1.0	15.0	36	0.40	0.5	3.0	16.0	14.8	0.0	0.0
8a:1730	0.0	15.0	116	0.30	0.5	3.6	16.0	15.1	0.0	0.0
8a:1830	1.0	15.0	0	0.1	0.6	3.9	16.1	15.1	3.6	0.0
8a:1930	1.0	15.3	0	0.1	0.5	3.6	16.0	14.6	3.9	0.0
8a:2030	1.5	15.0	0	0.1	0.0	3.0	16.0	14.7	3.9	0.0
8a:2130	1.0	15.0	0	0.1	1.7	0.9	16.0	14.8	3.8	0.0
8a:2230				0.1	0.0	3.9	16.0	14.8	3.8	
9a:0330	0.5	14.0	56	0.1	1.7	0.5	16.0	15.0	3.8	0.0
9a:0830	0.1	14.7	116	0.1	1.6	0.5	16.0	14.9	3.8	0.7
9a:0130	0.0	15.0	84	0.1	1.7	1.0	16.3	14.9	3.9	0.0
9a:0230	0.0	14.7	150	0.1	1.8	0.9	16.3	14.9	3.8	0.7
9a:0330	0.3	14.0	100	0.1	2.5	3.9	16.3	15.0	3.7	0.0
9a:0430	0.3	14.0	70	0.1	2.0	3.8	16.0	15.0	3.7	0.0
9a:0530	0.0	14.3	63	0.1	0.1	3.0	16.0	15.0	3.7	0.0
9a:0630	0.0	14.0	100	0.1	1.7	3.9	16.0	15.0	3.7	0.0
9a:0730	0.0	13.0	76	0.1	0.5	3.9	16.0	15.0	3.5	0.0
9a:0830	0.0	14.7	136	0.1	1.7	3.0	16.3	14.9	3.7	0.7
9a:0930	0.1	14.0	63	0.1	1.9	3.0	16.3	15.0	3.7	0.0
9a:1030	3.1	13.5	0	0.1	0.0	3.0	16.0	15.5	3.5	0.0
9a:1130	0.0	13.0	10	0.1	0.3	0.3	16.1	15.0	3.5	0.0
9a:1230	0.3	13.0	000	0.1	0.0	3.0	16.1	15.0	3.5	0.0
9a:1330	0.1	14.0	100	0.1	3.0	0.0	16.3	15.0	3.7	0.0
9a:1430	0.0	14.1	136	0.1	0.3	3.1	16.5	15.0	3.5	0.7
9a:1530	0.0	14.1	136	0.1	0.5	0.9	16.4	15.0	3.6	0.7
9a:1630	0.1	14.1	130	0.1	0.5	3.0	16.3	15.0	3.6	0.7
9a:1730	0.1	14.0	136	0.1	0.5	3.9	16.3	15.0	3.7	0.0
9a:1830	0.0	13.0	136	0.1	0.3	3.9	16.0	15.1	3.8	0.0
9a:1930	0.0	14.0	100	0.1	0.3	3.9	16.0	15.0	3.8	0.0
9a:2030	0.1	14.0	163	0.1	0.6	3.6	16.3	15.0	3.8	0.0
9a:2130	0.1	14.0	165	0.1	3.1	3.9	16.3	15.0	3.6	0.0
9a:2230	0.0	14.0	100	0.1	0.0	3.6	16.0	15.1	3.7	0.0
9a:2330	0.1	14.0	136	0.1	0.7	3.1	16.1	14.7	3.9	0.0
9a:0030	1.0	14.7	130	0.1	3.3	0.1	15.9	14.6	0.1	0.0
9a:0130	1.0	14.0	130	0.1	3.0	3.6	16.0	14.9	0.0	0.0
9a:0230	0.9	14.1	100	0.1	1.7	0.1	16.0	14.8	0.0	0.0
9a:0330	0.9	14.1	130	0.1	1.7	3.0	16.1	15.0	3.5	0.1
9a:0430	1.0	14.5	163	0.1	1.8	3.0	16.5	15.1	3.6	0.0
9a:0530	0.0	14.0	130	0.1	3.1	3.9	16.5	15.1	3.5	0.0
9a:0630	0.0	14.1	150	0.1	3.0	0.1	16.3	15.1	3.7	0.1

Table D-1

Run 1 Investigators Blank

CAFB GAS COMPOSITIONS

	FLUE GAS			REGENERATOR GAS			CATIFIER INLET GAS			
DAY/HOUR	O2 %	CO2 %	CO PPM	O2 %	CO2 %	CO PPM	VOL % ANAL	O2 CALC	VOL % ANAL	CO2 CALC
06/0730	1.6	18.0	165	.5	2.8	0.1	15.8	13.8	0.1	0.6
06/0830	1.7	17.1	189	.4	2.8	1.0	16.1	14.5	1.9	0.5
06/0930	2.0	18.5	135	.5	2.3	1.0	16.1	15.0	1.9	0.5
06/1030	2.0	18.5	135	.5	2.5	1.9	15.9	14.5	0.1	0.9
06/1130	2.3	18.6	100	.5	3.7	1.0	15.6	14.5	0.1	0.9
06/1230	2.0	18.1	135	.5	0.3	0.3	16.0	14.5	1.9	0.6
06/1330	2.0	18.1	135	.6	2.5	1.6	15.8	14.9	1.6	0.5
06/1430	2.0	18.0	189	.5	2.5	1.0	16.0	15.0	1.9	0.6
06/1530	2.0	18.1	163	.5	2.6	1.0	16.3	15.3	1.6	0.3
06/1630	2.0	18.0	000	0.50	2.5	1.0	16.3	15.3	1.7	0.3
06/1730	1.9	18.0	000	0.90	3.1	1.0	16.3	15.3	1.6	0.3
06/1830	2.0	18.1	007	.5	2.5	1.6	16.3	15.0	1.6	0.3

Table D-2

Run 3 Pennsylvania Blend

CAER PILOT PLANT PERFORMANCE

DAY:HOUR	SULFUR REMOVAL PCT	GAS RATE FT/SEC	U-BED DEPTH IN.	AIR/ FUEL % STOIC	CA/S MOL	REGENERATOR % CAC TO CAD	SULFUR OUTPUT LB/MBR	% OF FED
2:0130	85.8	4.98	97.1	93.0	2.07	38.9	3.33	34.6
2:0630	94.6	4.98	98.6	94.8	2.05	38.3	3.09	33.7
2:0930	75.9	4.96	97.1	92.7	2.75	40.2	3.52	37.9
2:0830	86.0	5.05	96.8	94.1	0.	33.5	2.99	31.9
2:0530	97.1	5.04	94.1	93.0	2.19	41.8	3.16	33.8
2:0630	97.9	4.93	94.3	93.8	2.19	36.8	3.08	34.1
2:0730	89.1	4.98	94.8	93.7	2.16	31.7	2.87	26.8
2:0930	88.6	5.07	96.7	94.3	2.07	21.8	1.86	19.8
2:1030	89.5	5.04	94.1	99.8	2.35	38.7	3.45	39.9
2:1130	85.0	5.09	97.1	98.6	2.61	28.8	2.33	23.6
2:1030	86.0	4.91	94.3	98.5	2.39	18.8	1.68	18.6
2:1330	81.7	4.91	94.7	91.8	2.86	18.5	1.56	16.8
2:1830	88.8	4.98	94.3	91.1	2.53	25.1	2.28	23.1
2:1530	88.6	4.76	97.1	98.8	2.55	39.7	3.18	30.8
2:1630	71.6	4.79	95.3	98.7	2.58	39.8	3.04	31.8
2:1730	76.5	4.84	96.8	98.7	2.56	36.5	3.28	38.8
2:1830	75.8	4.96	94.7	99.3	2.66	26.3	2.17	23.8
2:1930	77.6	4.96	94.3	99.3	2.66	28.8	2.17	23.6
2:0030	76.7	4.98	95.8	98.3	2.59	19.8	1.66	17.8
2:0130	77.6	4.56	95.3	98.8	2.59	18.6	1.59	16.8
2:0230	71.8	4.17	90.9	99.1	2.59	28.3	2.10	20.5
2:0330	75.6	4.09	90.9	96.1	2.66	38.9	2.33	25.1
3:0830	71.6	4.39	94.8	91.8	2.76	30.7	3.33	36.7
3:0130		4.71	95.8	99.8	2.68			
3:0230	71.6	4.03	95.8	98.8	2.65	25.8	2.08	22.1
3:0330	70.8	4.17	95.6	98.9	2.70	36.9	2.81	23.8
3:0430	73.8	4.01	94.3	98.5	2.68	36.6	2.86	20.8
3:0530	71.6	4.18	96.8	98.8	2.59	38.7	3.66	35.8
3:0630	74.6	4.38	94.3	98.3	2.50	38.9	3.7	35.6
3:0730	72.8	4.98	94.7	98.3	2.60	35.6	3.95	38.8
3:0830	78.8	4.83	94.7	98.8	2.60	35.3	3.98	38.3
3:0930	80.3	4.98	96.9	91.1	2.67	37.6	4.00	38.3
3:1030	73.8	4.11	95.3	98.8	2.58	35.6	3.39	35.1
3:1130	76.5	4.58	94.7	98.9	2.63	17.8	1.80	18.8
3:1030	75.9	4.58	95.3	91.8	2.63	07.8	0.85	09.9
3:1330	78.8	4.30	91.8	19.9	2.66	0.8	0.81	0.3
3:1830	83.1	4.38	91.5	98.8	2.58	48.6	4.16	43.8
3:1530	73.8	4.48	97.5	98.7	2.53	27.8	2.67	28.8
3:1630	75.3	4.66	96.7	98.9	2.68	35.8	3.67	37.3
3:1730	67.8	4.53	93.7	91.8	2.61	31.7	2.68	28.8

Table D-2

Run 2 Demineralizer Runs

CAER PILOT PLANT PERFORMANCE

DAY/HOUR	SULFUR REMOVAL PCT	GAS RATE FT/SEC	SCRUB DEPTH IN.	AIR/ FUEL # STOIC	CA/S MOL	% CAS TO FGD	REGENERATION SULFUR OUTPUT LB/HR	% OF FED
1.1530	86.0	8.51	35.7	19.8	0.66	36.1	1.73	36.7
1.1930	58.0	8.53	80.5	20.0	0.63	36.5	1.68	38.3
1.2030	58.9	8.65	80.5	18.3	0.57	8.8	1.03	8.6
1.2130	69.1	8.65	80.5	20.6	0.69	10.0	0.75	7.8
1.2530	69.5	8.51	80.5	20.8	0.63	11.6	1.65	39.7
1.2730		8.63	80.0	20.3	0.	11.6	1.57	13.0
1.2830		8.67	80.5	21.3	0.	09.1	1.66	11.6
2.0130	71.6	8.16	80.5	19.8	0.	32.6	0.12	60.1
2.0230	71.6	8.81	80.5	20.0	0.	85.7	5.18	50.6
2.0330	70.9	8.17	31.9	20.9	0.	36.1	0.85	63.6
2.0430	70.8	8.25	31.9	20.5	0.	36.5	0.88	60.6
22.2030	95.1	8.61	86.6	20.1	0.30	04.0	0.55	04.6
22.2130	96.6	8.36	86.7	20.6	1.53	10.6	1.65	1.5
22.2230	95.5	8.33	89.3	20.0	1.71	09.1	1.35	10.0
22.2330	93.1	8.66	86.7	20.6	1.71	11.1	1.86	14.5
23.0030	96.0	8.33	89.3	20.8	1.65	10.1	1.75	13.5
23.0130	98.6	8.09	87.3	20.1	1.60	10.0	1.66	14.0
23.0230	99.9	8.36	12.0	20.9	1.57	04.0	0.75	06.3
23.0330	97.1	8.07	12.0	19.1	1.12	06.0	1.16	07.1
23.0630	99.3	8.05	11.1	20.0	1.16	11.0	1.70	13.1
23.0530	99.5	8.04	27.5	20.0	1.16	10.1	0.67	10.0
23.0630	97.6	8.09	30.1	20.1	1.10	11.1	1.71	12.9
23.0730	98.9	8.09	04.1	19.8	1.06	09.0	1.05	08.9
23.0830	98.9	8.09	87.5	19.0	1.08	10.0	0.11	15.0
23.0930	98.9	8.09	30.1	19.5	1.09	11.6	1.70	13.0
23.1030	97.9	8.09	38.8	19.5	1.01	10.0	1.30	09.1
23.1130	97.7	8.03	24.5	17.1	1.06	15.5	1.70	13.6
23.1230	96.5	8.03	30.8	18.9	1.04	09.6	0.86	15.0
23.1330	98.0	8.06	87.5	19.0	1.03	14.6	1.90	14.5
23.1430	91.9	8.05	87.5	19.1	1.15	19.0	0.80	15.6
23.1530	94.6	8.01	30.7	19.9	1.05	10.1	1.06	08.1
23.1630	97.6	8.05	30.0	19.6	1.01	55.0	0.75	61.6
23.1730	98.5	8.03	30.8	19.8	1.05	10.3	1.70	13.5
23.1830	97.1	8.07	30.7	19.8	1.06	10.3	0.57	08.3
23.1930	97.3	8.08	30.1	19.8	1.06	12.0	0.81	08.5
23.2030	98.1	8.06	86.7	19.8	1.06	15.0	1.65	14.4
23.2130	96.5	8.08	87.5	19.9	0.78	10.1	1.31	13.5
23.2230	94.0	8.05	86.3	19.8	0.98	10.8	1.09	11.3
23.2330	93.3	8.05	86.9	19.0	1.09	10.7	1.01	10.6
24.0030	96.6	8.06	89.3	19.5	1.06	10.1	1.05	10.3

Table D-2

Run 3 Pennsylvania Stone

CAFB PILOT PLANT PERFORMANCE

DAY:HOUR	SULFUR REMOVAL PCT	GAS RATE FT/SEC	U-BED DEPTH IN:	AIR/ FUEL % STOIC	CA/S MOL	REGENERATOR % GAS TO CAD	SULPHUR OUTPUT LB/HR	% OF FED
00:0130	93.5	4.84	08.7	19.3	1.65	36.9	3.98	38.7
00:0230	94.6	4.94	10.0	19.7	1.80	38.0	3.98	39.6
00:0330	87.1	4.84	10.7	19.5	1.18	65.0	4.15	41.7
00:0430	89.5	4.84	08.7	19.1	0.96	38.0	3.64	35.7
00:0530	90.5	4.84	08.7	19.0	0.93	37.6	4.04	39.6
00:0630	90.5	4.85	08.7	19.6	1.04	41.6	4.31	44.0
00:0730	91.7	3.95	08.7	00.0	1.04	38.5	3.70	36.1
00:0830	93.9	4.01	10.7	19.1	1.19	46.5	4.65	45.5
00:0930	93.6	4.06	08.7	19.4	1.15	38.1	3.86	38.4
00:1030	98.9	4.80	04.7	19.5	1.00	84.5	8.58	85.5
00:1130	89.3	4.83	10.7	19.6	0.93	31.7	3.31	33.4
00:1230	88.1	4.83	09.6	19.7	1.02	40.6	3.86	38.8
00:1330	83.7	4.84	09.6	19.8	1.04	44.3	4.55	44.7
00:1430	78.0	4.80	10.0	19.8	1.03	41.3	4.01	40.5
00:1530	83.8	4.84	10.0	19.8	0.93	38.0	4.01	40.5
00:1630	81.0	4.84	10.0	19.9	0.96	38.0	3.81	38.6
00:1730	87.3	4.83	10.0	19.7	1.16	40.0	4.06	40.9
00:1830	85.0	4.85	10.0	19.8	0.96	36.8	3.61	36.6
00:1930	84.1	4.83	10.0	19.7	1.13	13.5	1.69	15.1
00:2030	78.8	4.81	06.6	17.8	0.37	04.9	0.76	08.0
00:2130	88.7	4.87	08.6	18.6	0.37	04.9	0.76	08.5
00:2230	95.1	4.87	10.0	18.8	0.63	33.7	3.78	38.8
00:2330	100.0	4.91	10.7	19.7	2.03	9.5	1.03	10.8
01:0030	99.3	1.84	11.6	17.8	0.81	03.8	0.38	03.9
01:0130	99.1	1.99	10.9	18.7	0.10	00.3	0.09	11.1
01:0230	99.1	4.91	10.9	18.9	0.33			
01:0330	99.1	4.10	11.6	00.0	0.31	35.3	3.86	40.1
01:0430	99.0	4.88	10.9	19.8	0.98	08.9	1.18	10.6
01:0530	99.0	4.10	10.9	19.5	0.30	33.3	3.61	36.8
01:0630	99.3	4.10	10.9	19.6	0.10	14.3	1.57	16.0
01:0730	98.8	4.11	14.3	19.8	0.04	31.4	3.38	33.3
01:0830	99.0	4.10	10.9	19.5	0.14	00.1	0.00	00.0
01:0930	99.6	4.09	10.9	19.5	0.00	09.3	1.19	10.1
01:1030	99.6	4.08	10.9	18.1	0.04	41.6	4.30	43.3
01:1130	99.6	1.93	10.9	19.8	0.58	36.8	3.99	38.4
01:1230	99.6	4.09	10.9	19.9	0.09	04.8	0.52	05.5
01:1330	99.6	4.10	15.6	19.9	0.09	04.8	0.50	04.1
01:1430	99.6	4.11	10.6	01.0	0.09	16.6	1.67	17.9
01:1530	99.6	4.08	10.6	19.7	0.58	36.3	3.33	36.1
01:1630	99.6	4.01	15.7	19.7	1.93	38.5	4.08	41.1
01:1730	99.6	4.15	14.6	19.8	0.67	18.5	1.76	18.8
01:1830	99.6	4.08	13.3	19.9	0.94	34.8		

Table IV-1

Run 3 Pennsylvania Blend

CAFB PILOT PLANT PERFORMANCE

DAY:HOUR	SULPHUR REMOVAL PCT	GAS RATE FT/SEC	BURNED DEPTH IN.	AIR/ FUEL % STOIC	CA/S MOL	REGENERATOR % CAS TO CAO	SULPHUR OUTPUT LB/HR	% OF FED
0A:1530	96.4	4.00	10.7	18.8	1.38	16.6	1.78	17.8
0A:1830	97.6	4.07	10.7	18.8	1.39	10.8	1.85	18.9
0A:2130	97.3	4.20	10.3	19.7	1.51	11.7	1.80	18.5
0A:2430	[98.0	3.99	15.7	18.7	1.47	16.5	1.70	17.3
0A:2730	[98.0	4.04	15.7	19.6	1.68	11.8	1.95	18.0
0A:3030	[98.0	4.04	15.7	18.7	1.55	07.7	1.19	11.6
0A:3330	[98.0	4.09	15.7	18.9	1.58	08.3	0.70	09.7
0A:0030		4.09	17.1	18.0	1.68	11.5	1.89	19.8
0A:0330	96.3	4.11	15.7	19.0	1.46	01.0	0.17	5.1
0A:0630	93.1	4.04	15.7	19.8	1.68	08.8	1.55	05.9
0A:0930	94.7	4.10	15.7	19.8	1.75	15.7	1.76	17.9
0A:1230	96.8	4.08	15.7	19.8	1.68	03.9	0.73	07.8
0A:1530	93.9	4.10	16.8	19.1	1.68	16.1	1.96	18.8
0A:1830	95.1	4.10	16.8	19.3	1.70	13.0	1.67	17.5
0A:2130	95.7	4.09	15.7	19.3	1.68	10.8	1.66	17.6
0A:2430	94.3	4.10	15.7	19.0	1.59	10.7	1.80	19.0
0A:2730	96.6	4.13	15.7	19.6	1.77	16.0	1.96	19.3
0A:3030	97.9	4.09	15.7	19.0	1.60	08.6	1.11	10.1
0A:3330	95.7	1.58	16.5	18.8	1.67	07.6	1.13	10.1
0A:0030	[98.3	4.09	16.1	19.6	1.59	10.7	1.51	15.8
1A:0330	96.7	1.98	16.1	18.5	1.07	16.5	0.16	00.7
0A:0630	85.9	1.85	16.7	18.2	1.53	13.8	1.17	15.6
0A:0930	91.7	4.01	16.6	19.1	1.67	01.3	0.08	05.5
0A:1230	93.6	4.01	16.6	19.3	1.76	08.8	1.08	07.3
0A:1530	90.6	4.00	16.6	19.8	1.91	05.3	0.78	06.0
0A:1830	86.3	4.03	03.3	19.5	1.41	09.9	1.09	13.9
0A:2130	98.9	1.99	17.1	18.7	0.96	13.5	1.03	18.1
0A:2430	90.6	4.09	16.6	19.1	1.86	13.0	1.01	10.5
0A:2730	97.1	4.00	16.5	18.9	1.90	13.9	1.71	17.3
0A:3030	86.9	4.01	16.5	18.8	1.15	11.2	1.16	16.1
0A:3330	90.1	4.10	17.7	19.0	0.98	15.5	1.70	17.3
0A:0030	91.6	4.11	17.3	19.3	1.13	14.6	1.58	10.1
0A:0330	90.6	1.96	17.3	18.8	1.16	08.8	1.08	10.3
0A:0630	91.3	1.93	15.7	17.0	1.16	10.7	0.91	09.5
0A:0930	93.0	4.06	15.7	18.8	1.03	10.0	1.63	16.5
0A:1230	91.5	1.98	15.7	18.8	1.10	09.7	1.56	25.7
0A:1530	90.3	4.08	15.7	19.3	1.83	07.8	1.14	11.9
0A:1830	86.0	4.03	15.7	18.7	1.09	07.5	1.16	11.9
0A:2130	91.8	4.03	15.7	18.8	1.81	15.7	1.90	19.5
0A:2430	89.1	4.03	15.7	18.8	0.87	17.8	0.98	09.5

Table No.1

Run 5 Pennsylvania Blend

CARB PILOT PLANT PERFORMANCE

DAY:HOUR	SULPHUR REMOVAL PCT	GAS RATE FT/SEC	C-REC DEPTH IN.	BED FUEL % STOIC	GAZE RQL	DEFLUIDATOR % CAR TO CAR	SULPHUR OUTPUT LB/MM	% OF FED
06:0730	90.1	6.04	35.7	18.7	1.38	17.9	1.99	88.1
06:0830	90.6	3.95	35.7	18.3	1.41	86.9	3.01	87.5
06:0930	90.7	6.00	35.7	18.8	1.03	98.6	3.33	93.6
06:1030	90.8	3.61	38.5	17.6	1.37	34.1	3.00	89.0
06:1130	93.0	3.80	37.7	17.5	1.41	38.6	3.07	93.0
06:1230	89.6	3.35	38.5	17.0	1.66	69.9	6.33	83.6
06:1330	95.6	3.80	37.7	17.9	1.35	31.3	3.56	95.9
06:1630	90.6	6.00	35.5	18.5	1.66	98.3	3.18	97.1
06:1530	88.7	3.33	39.0	18.6	1.67	30.1	3.33	93.6
06:1630	84.6	3.93	39.0	18.7	1.38	98.7	3.13	91.5
06:1730	84.6	3.96	38.7	18.3	1.35	31.6	3.70	93.0
06:1830	84.1	3.85	35.5	18.9	1.36	31.3	3.68	96.1

Table [illegible]

[illegible] [illegible] [illegible] [illegible]

[illegible] [illegible] [illegible]

[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]

Table D-[illegible]

Run [illegible]

[illegible] GAS COMPOSITIONS

	FLUE GAS			REGENERATOR GAS			[illegible] INLET GAS			
DAY,HOUR	[illegible] %	[illegible] %	[illegible] PPM	O2 %	[illegible] %	[illegible] %	[illegible]	[illegible]	[illegible]	[illegible]
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]

Table No.2

[illegible]

[illegible]

Table [illegible]

[illegible] Unit [illegible]

[illegible] GAS COMPOSITIONS

[illegible]	BLOW GAS			SECONDARY GAS			CALCULATED INLET GAS			
	O2	CO2	CO	O2	CO2	CO	O2	% O2	CO2	% CO2
[illegible]		%	%	%		%	ANAL	CALC	ANAL	CALC
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]

[illegible]

[illegible]

Table IV-5

Run #68 1011 Blend

COED PILOT PLANT PERFORMANCE

Days	Sulphur Removal %	Gas Rate SCFH/Ft²	Fixed Depth In.	Air/Fuel # Steam/# Coal	Coal Rate #/Hr/Ft²	Regeneration % Gas to Gas Loops	Regeneration Sulphur Output Lb/Hr	Regeneration # of Fed
1.0-10.0	77.1	6.33	20.1	28.6	0.91	33.1	1.16	13.3
1.0-11.0	83.9	6.33	25.6	28.1	1.39	68.0	1.59	19.8
1.0-12.0	88.7	6.30	25.6	19.9	1.30	79.6	2.75	29.5
1.0-13.0	93.0	6.25	26.1	28.7	1.30	79.6	3.27	35.6
1.0-14.0	93.0	6.06	27.1	28.1	1.13	86.9	4.05	43.7
1.0-15.0	92.6	6.06	25.1	28.8	1.10	89.5	3.60	39.9
1.0-16.0	88.0	5.70	27.1	[illegible]	1.00	53.5	0.76	53.3
1.0-17.0	75.3	5.66	21.6	28.8	1.53			
1.0-18.0	75.6	6.58	26.1	21.6	1.56	5.1	0.52	6.2
1.0-19.0	71.8	6.58	25.1	28.1	0.80			
1.0-20.0	79.6	6.35	26.0	23.0	0.93	13.1	1.56	17.3
1.0-21.0	82.7	6.01	26.1	21.1	1.01	13.6	1.49	18.3
1.0-22.0	81.6	5.86	26.6	21.6	1.60	0.1	0.01	0.1
1.0-23.0	69.3	6.03	25.0	21.5	1.19	0.0	0.00	0.0
1.0-25.0	81.5	6.03	26.1	21.1	0.93	0.1	0.01	0.0
11.1-12.1	78.3	6.57	21.7	21.6	0.59	23.6	2.80	33.0
11.1-13.0	83.1	6.30	25.6	25.1	1.11	17.1	1.06	17.5
11.1-14.0	84.6	6.33	27.0	21.5	0.98	09.0	1.15	17.3
11.1-15.0	81.9	6.33	26.1	21.6	0.98	19.1	2.33	33.0
11.1-16.0	83.3	6.30	26.1	21.6	1.06	3.6	0.59	6.5
11.1-17.0	78.0	6.03	21.6	25.0	1.06	09.1	1.66	26.1
11.1-18.0	78.0	6.30	21.6	20.0	1.01	00.0	1.56	25.0
11.1-19.0	81.0	6.15	21.6	20.1	1.01	13.1	0.75	8.0
11.1-20.0	82.9	6.36	26.1	21.1	1.16	07.1	0.93	11.1
11.1-21.0	79.1	6.30	27.1	21.1	1.05	03.1	1.00	15.0
11.1-22.0	78.3	6.07	26.1	21.6	1.07	04.1	1.08	16.9
11.1-23.0	81.0	6.33	26.0	22.1	1.03	16.0	1.58	24.1
11.1-24.0	79.0	6.36	21.6	21.1	1.16	05.1	1.03	16.1
11.1-25.0	78.0	6.36	21.6	20.1	1.16	05.1	1.03	16.1
11.1-26.0	78.0	6.36	21.6	20.1	1.16	05.1	1.03	16.1
11.1-27.0	78.0	6.36	21.6	20.1	1.16	05.1	1.03	16.1
11.1-28.0	78.0	6.36	21.6	20.1	1.16	05.1	1.03	16.1
11.1-29.0	78.0	6.36	21.6	20.1	1.16	05.1	1.03	16.1
11.1-30.0	78.0	6.36	21.6	20.1	1.16	05.1	1.03	16.1
11.1-31.0	78.0	6.36	21.6	20.1	1.16	05.1	1.03	16.1
11.1-32.0	78.0	6.36	21.6	20.1	1.16	05.1	1.03	16.1
11.1-33.0	78.0	6.36	21.6	20.1	1.16	05.1	1.03	16.1
11.1-34.0	78.0	6.36	21.6	20.1	1.16	05.1	1.03	16.1
11.1-35.0	78.0	6.36	21.6	20.1	1.16	05.1	1.03	16.1
11.1-36.0	78.0	6.36	21.6	20.1	1.16	05.1	1.03	16.1
11.1-37.0	78.0	6.36	21.6	20.1	1.16	05.1	1.03	16.1
11.1-38.0	78.0	6.36	21.6	20.1	1.16	05.1	1.03	16.1
11.1-39.0	78.0	6.36	21.6	20.1	1.16	05.1	1.03	16.1
11.1-40.0	78.0	6.36	21.6	20.1	1.16	05.1	1.03	16.1
11.1-41.0	78.0	6.36	21.6	20.1	1.16	05.1	1.03	16.1
11.1-42.0	78.0	6.36	21.6	20.1	1.16	05.1	1.03	16.1
11.1-43.0	78.0	6.36	21.6	20.1	1.16	05.1	1.03	16.1
11.1-44.0	78.0	6.36	21.6	20.1	1.16	05.1	1.03	16.1
11.1-45.0	78.0	6.36	21.6	20.1	1.16	05.1	1.03	16.1
11.1-46.0	78.0	6.36	21.6	20.1	1.16	05.1	1.03	16.1
11.1-47.0	78.0	6.36	21.6	20.1	1.16	05.1	1.03	16.1
11.1-48.0	78.0	6.36	21.6	20.1	1.16	05.1	1.03	16.1
11.1-49.0	78.0	6.36	21.6	20.1	1.16	05.1	1.03	16.1
11.1-50.0	78.0	6.36	21.6	20.1	1.16	05.1	1.03	16.1

Table D-5

Run 5 BWR High Burnup

CASE PILOT PLANT CHARACTERISTICS

CHARACTERISTIC	CALENDAR DESIGNATION	UNIT	CORE DIAMETER (EQUIVALENT)	CORE HEIGHT (EQUIVALENT)	AREA FUEL	ZONES	SPECIFICATIONS	
							# ZONES	ENRICHMENT (WT%)
							TO END LOADS	# OF ZONES

Table D-5

Run 5 ROR Test Block

COED PILOT PLANT PERFORMANCE

DAY:HOUR	SULPHUR REMOVAL PCT	GAS RATE FT/SEC	COKED DEPTH IN	AIR/ FUEL % STOIC	GAZA RUN	% GAS TO GAS LEVNO	REGENERATOR SULPHUR OUTPUT % OF FED
17:1030	88.0	0.94	38.5	21.7	1.35	37.8	83.1
17:1130	89.5	0.16	15.7	90.7	1.80	88.0	91.99	91.7
17:1230	91.1	0.18	15.7	21.8	1.80	87.7	92.88	88.0
17:1330	93.9	0.95	15.7	20.9	1.19	89.0	91.0	89.9
17:1430	71.7	0.35	15.7	21.5	1.31	87.5	91.85	80.7
17:1530	74.5	0.36	15.7	20.5	1.58	11.3	92.0	63.8
17:1630	96.0	0.17	15.7	21.8	1.58	20.8	92.86	95.8
17:1730	73.0	0.58	11.1	20.3	1.31	12.0	91.79	72.5
17:1830	91.5	0.17	11.1	21.8	1.17	88.5	91.98	91.5
17:1930	0.0	0.15	11.1	21.3	1.09	33.1	91.05	88.0
17:2030	91.0	0.15	11.1	21.0	0.98	89.5	91.16	75.3

Table [illegible]

Run 3 Temperatures, Fuel and Slope Rates

DAY:TIME	TEMPERATURE: DEG F:				RATE: LB/HR
	BASIFIER	OF-EN	RECYCLE	OIL	STEAM
1:0130	933:	1061:	72:	604:5	71:0
1:0230	923:	1036:	70:	603:6	63:0
1:0330	939:	1069:	75:	606:0	69:0
1:0430	948:	1068:	76:	605:0	0:
1:0530	948:	1038:	76:	597:1	5:5
1:0630	949:	1053:	76:	597:1	5:5
1:0730	933:	1058:	73:	604:3	8:0
1:0930	948:	1060:	75:	599:6	6:0
1:1030	959:	1055:	75:	614:5	11:0
1:1130	965:	1070:	75:	614:5	13:0
1:1330	968:	1068:	79:	607:0	12:0
1:1430	965:	1059:	75:	607:6	12:0
1:1630	966:	1059:	75:	598:0	10:0
1:1730	969:	1058:	80:	605:6	11:0
1:1830	965:	1058:	80:	603:0	10:0
1:1930	963:	1058:	80:	606:6	11:0
1:2030	968:	1060:	75:	597:0	08:0
1:2130	968:	1065:	75:	597:0	08:0
1:2230	960:	1030:	69:	599:0	10:0
1:2330	965:	1061:	65:	598:0	10:0
2:0030	973:	1060:	65:	598:1	10:0
2:0130	969:	1065:	69:	598:0	08:0
2:0230	969:	1059:	75:	599:0	08:0
2:0330	969:		71:	603:5	11:0
2:0430	965:	1061:	75:	607:0	08:0
2:0530	964:	1065:	69:	605:5	11:0
2:0630	969:	1063:	69:	599:0	10:0
2:0730	969:	1063:	69:	599:0	10:0
2:0830	969:	1057:	75:	599:0	10:0
2:0930	971:	1058:	75:	601:1	10:0
2:1030	969:	1059:	70:	597:0	08:0
2:1130	971:	1061:	70:	597:0	10:0
2:1230	971:	1055:	70:	599:0	11:0
2:1330	970:	1065:	70:	601:5	13:0
2:1430	967:	1060:	70:	603:0	10:0
2:1530	970:	1070:	70:	598:1	14:5
2:1630	971:	1061:	70:	596:5	15:0
2:1730	969:	1065:	69:	596:5	10:0
2:1830	968:	1061:	70:	594:0	11:0
2:1930	970:	1060:	70:	600:0	11:0
2:2030	970:	1065:	75:	595:0	10:0
2:2130	975:	1065:	70:	634:0	10:0
2:2230	979:	1070:	69:	603:0	11:0
2:2330	980:	1065:	65:	598:1	11:0
3:0030	980:	1067:	65:	597:0	0:

Table D-5

Run 3 Temperatures, Fuel and Stone Rates

DAY:TIME	TEMPERATURE, DEG F:			RATE, LB/HR	
	GASIFIER	BEDEN	RECYCLE	OIL	STONE
5:08:30	901	1862	53	379.8	0
5:09:30	876	1860	69	868.5	0
5:10:30	931	1855	89	883.2	0
5:11:30	886	1860	80	385.3	0
5:12:30	910	1865	83	391.7	0
5:13:30	850	1867	55	399.9	101.0
5:14:30	838	1866	65	800.6	136.0
5:15:30	853	1866	105	399.9	110.0
5:16:30	850	1863	80	801.7	57.0
5:17:30	870	1858	78	399.6	19.0
5:18:30	878	1858	75	397.0	0
5:19:30	869	1873	70	399.0	0
5:20:30	875	1868	79	393.5	84.0
5:21:30	870	1858	68	396.0	64.0
5:22:30	861	1858	68	397.6	63.0
5:23:30	860	1853	68	396.0	65.0
5:24:30	858	1868	58	398.0	63.0
5:25:30	856	1858	60	395.3	61.0
5:26:30	868	1858	66	396.8	70.0
5:27:30	867	1858	68	397.0	65.0
5:28:30	868	1858	68	398.0	70.0
5:29:30	869	1854	86	393.5	59.0
5:30:30	865	1853	85	393.5	60.0
6:00:30	873	1858	68	398.3	63.0
6:00:43	868	1855	75	413.6	73.0
6:01:30	870	1855	68	398.0	65.0
6:02:30	867	1858	68	378.3	63.0
6:03:30	865	1858	63	387.1	64.0
6:04:30	858	1858	68	413.6	67.0
6:05:30	858	1853	85	408.6	68.0
6:06:30	858	1858	68	408.6	63.0
6:07:30	858	1858	68	399.0	58.0
6:08:30	858	1858	68	408.6	65.0
6:09:30	865	1858	68	408.6	64.0
7:13:30	875	1858	68	399.0	67.5
7:11:30	874	1858	68	399.0	68.0
7:12:30	865	1849	68	398.1	60.0
7:13:30	870	1855	68	408.0	70.0
7:14:30	870	1853	68	406.0	69.0
7:15:30	888	1858	68	406.0	75.0
8:00:30	865	1849	85	388.7	70.0
8:01:30	860	1858	83	388.7	69.0
8:02:30	859	1856	88	388.3	67.0
8:03:30	868	1858	85	381.6	66.0
8:04:30	858	1868	83	388.5	60.0
8:05:30	858	1855	83	381.6	70.0
8:06:30	858	1865	88	387.9	61.0

Table 3c.3

Run 3 Temperatures, Fuel and Stone Rates

DAY:TIME	TEMPERATURE: DEG C:			RATE: LB/HR	
	CASIFIER	DEGEN	RECYCLE	OIL	STONE
8:0030	895:	1005:	30:	1011.6	58.0
8:0130	890:	1015:	30:	1004.7	55.8
8:0230	870:	1003:	30:	1011.6	71.0
8:0330	870:	1040:	30:	1011.5	70.0
8:0430	875:	1040:	30:	1002.6	65.0
8:0530	870:	1040:	30:	1011.5	59.0
8:0630	880:	1040:	30:	1011.7	63.0
8:0730	880:	1055:	30:	1011.5	70.0
11:1230	871:	1040:	60:	1002.6	87.0
11:1530	860:	1060:	60:	1011.6	64.0
11:1830	860:	1060:	60:	1011.5	75.0
11:1130	864:	1064:	60:	1011.5	75.0
11:1030	870:	1060:	70:	1002.6	70.0
11:1330	880:	1065:	60:	1002.6	76.0
11:0030	890:	1075:	60:	1011.5	70.0
11:0130	890:	1067:	60:	1011.7	84.0
11:0230	880:	1060:	85:	1002.1	87.0
11:0330	867:	1050:	60:	1002.1	86.0
10:0030	869:	1065:	65:	887.9	51.0
10:0230	875:	1065:	65:	889.1	70.0
10:0130	881:	1065:	60:	887.5	78.0
10:0830	860:	1050:	60:	878.9	94.0
10:0530	851:	1050:	60:	887.5	91.0
10:0630	870:	1050:	60:	887.5	76.0
10:0730	875:	1050:	75:	1011.5	74.0
10:0830	870:	1050:	75:	1004.1	78.0
12:0930	860:	1060:	75:	1011.1	69.0
10:1030	865:	1060:	60:	1002.6	85.0
10:1130	875:	1070:	60:	1015.1	69.0
10:1430	860:	1060:	60:	1002.0	81.0
10:1630	860:	1050:	60:	1002.0	81.5
12:1530	870:	1060:	75:	1015.1	87.0
10:1630	860:	1071:	83:	1011.0	85.0
12:1730	865:	1060:	60:	1015.1	78.0
10:1830	870:	1071:	85:	1011.5	91.0
10:1930	875:	1085:	85:	1002.6	87.0
10:2030	870:	1070:	85:	1002.6	76.0
10:2130	871:	1070:	81:	1002.6	81.5
10:2030	870:	1075:	80:	1011.0	79.5
10:2130	875:	1060:	75:	1002.1	75.0
11:0030	865:	1060:	75:	1002.1	69.0
11:0130	867:	1065:	75:	1002.0	80.0
11:0230	870:	1060:	75:	1011.1	78.5
11:0330	875:	1060:	80:	1011.5	78.0
11:0430	880:	1060:	80:	1011.1	78.5
11:0530	880:	1065:	75:	1002.0	86.0
11:0630	870:	1060:	75:	1011.1	71.0

Table D-3

Run 3 Temperatures, Fuel and Stone Rates

DAY:TIME	TEMPERATURE; DEG C;			RATE; LB/HR	
	CASIFIER	REFUEN	RECYCLE	OIL	STONE
15:07:10	868;	998;	08;	387;1	39;0
15:08:10	875;	1070;	81;	387;1	56;0
15:09:10	846;	1068;	81;	388;9	55;0
15:10:10	869;	1070;	80;	388;0	49;0
15:11:10	881;	1069;	80;	387;1	45;0
15:12:10	866;	1050;	79;	388;9	61;0
15:13:10	869;	1051;	79;	388;3	63;0
15:15:10	870;	1060;	79;	388;7	68;0
15:16:10	861;	1068;	71;	361;6	65;0
15:17:10	865;	1068;	71;	361;6	70;0
15:18:10	870;	1061;	71;	381;6	77;0
15:19:10	865;	1060;	70;	381;5	75;0
15:20:10	859;	1055;	70;	379;8	73;0
15:21:10	855;	1055;	70;	380;8	73;0
15:22:10	867;	1060;	70;	380;6	73;0
15:23:10	867;	1058;	70;	380;6	74;0
16:00:10	861;	1061;	70;	396;3	74;0
16:01:10	870;	1060;	70;	397;3	77;0
16:02:10	870;	1059;	70;	399;0	73;0
16:03:10	870;	1061;	70;	397;3	74;0
16:04:10	870;	1060;	70;	399;0	80;0
16:05:10	870;	1060;	70;	410;2	80;0
16:06:10	876;	1059;	69;	376;9	85;0
16:07:10	870;	1060;	70;	399;0	80;0
17:07:10	869;	1059;	70;	397;3	81;0
17:08:10	876;	1058;	69;	399;0	86;0
17:09:10	869;	1057;	70;	399;0	85;3
17:10:10	861;	1058;	70;	397;3	86;0
17:11:10	860;	1059;	70;	399;0	87;0
17:12:10	861;	1058;	70;	400;6	87;0
17:13:10	862;	1058;	70;	399;0	88;0
17:14:10	860;	1058;	70;	411;8	90;0
17:15:10	865;	1061;	70;	396;3	89;0
17:16:10	858;	1061;	70;	398;1	86;0
17:17:10	860;	1063;	70;	397;3	89;0
17:18:10	858;	1058;	70;	397;3	83;0
17:19:10	869;	1061;	65;	397;3	80;0
17:20:10	860;	1058;	69;	397;3	86;0
20:06:10	910;	1013;	65;	406;3	86;0
20:21:10	930;	1013;	60;	410;3	91;0
20:22:10	889;	1055;	70;	406;8	89;0

Table D-5

Run 3 Temperatures, Fuel and Blend Rates

DAY:TIME	TEMPERATURE: DEG F:			RATE: LB/HR	
	DAVIFIER	BEDEX	RECYCLE	OIL	STONE
08:0130	898:	1052:	10:	869:1	60:0
08:0230	915:	1055:	65:	869:1	57:0
08:0130	915:	1053:	65:	870:1	53:0
08:0230	915:	1060:	68:	871:6	56:0
08:0330	918:	1056:	65:	877:7	80:0
08:0630	905:	1060:	65:	871:3	60:0
08:0530	900:	1060:	65:	871:3	60:0
08:0630	910:	1060:	65:	869:5	59:0
08:0730	900:	1060:	65:	869:5	65:0
08:0830	896:	1060:	60:	877:7	55:0
08:0930	901:	1061:	65:	859:8	38:0
08:1030	900:	1060:	65:	869:5	61:0
08:1130	899:	1061:	65:	867:6	65:0
08:1230	900	1060:	67:	865:5	66:0
08:1330	903:	1059:	64:	867:5	51:0
08:1430	901:	1060:	67:	867:6	60:0
08:1530	899:	1060:	66:	870:5	50:0
08:1630	905:	1061:	63:	870:8	51:0
08:1730	896:	1060:	65:	870:5	50:0
08:1830	875:	1060:	66:	865:6	51:0
08:1930	893:	1060:	68:	870:8	55:0
08:2030	890:	1060:	70:	875:7	35:0
08:2130	868:	1060:	70:	869:1	34:0
08:2230	856:	1060:	70:	871:3	30:0
08:2330	870:	1071:	70:	872:6	60:0
09:0030	871:	1065:	70:	870:9	60:0
09:0130	851:	1070:	70:	870:7	66:0
09:0230	878:	1060:	70:	869:1	31:0
09:0330	870:	1069:	70:	871:8	37:0
09:0430	870:	1070:	70:	869:5	30:0
09:0530	875:	1062:	70:	869:7	30:0
09:0630	875:	1070:	70:	869:8	30:0
09:0730	860:	1070:	70:	870:7	30:0
09:0830	875:	1070:	70:	869:5	36:0
09:0930	870:	1071:	70:	871:7	36:0
09:1030	870:	1069:	70:	870:9	30:0
09:1130	871:	1069:	70:	870:8	29:0
09:1230	875:	1065:	70:	870:9	30:0
09:1330	873:	1069:	70:	870:8	30:0
09:1430	873:	1070:	70:	869:1	29:0
09:1530	870:	1069:	71:	869:1	30:0
09:1630	871:	1069:	70:	869:1	30:0
09:1730	871:	1071:	71:	870:9	30:0
09:1830	873:	1068:	70:	869:1	30:0

Table No1

Run 1 Temperatures, Fuel and Bleed Rates

DAY:TIME	TEMPERATURE: DEG F:			RATE: LB/HR	
	GASIFIER	BEDEN	CYCLOLE	OIL	BLEED
03:1730	891:	1053:	751	393:5	71:0
03:1830	869:	1055:	751	404:5	73:0
03:1930	868:	1050:	751	395:3	71:0
03:2030	869:	1061:	761	398:6	67:0
03:2130	868:	1061:	751	398:1	67:0
03:2230	859:	1060:	751	396:0	70:0
03:2330	865:	1060:	751	395:1	70:0
04:0030	869:	1060:	751	399:1	70:0
04:0130	863:	1060:	751	402:6	70:0
04:0230	867:	1060:	751	405:0	71:0
04:0330	855:	1060:	751	409:1	66:0
04:0430	867:	1055:	751	409:1	64:0
04:0530	860:	1059:	761	410:3	67:0
04:0630	861:	1059:	751	410:0	63:0
04:0730	865:	1061:	751	410:7	66:0
04:0830	870:	1061:	761	410:0	75:0
04:0930	873:	1061:	761	411:0	70:0
04:1030	873:	1061:	761	409:1	69:0
04:1130	875:	1060:	751	401:7	71:0
04:1230	873:	1061:	771	409:1	69:0
04:1330	875:	1061:	791	410:0	71:0
04:1430	870:	1060:	781	409:1	70:0
04:1530	874:	1060:	781	410:0	70:0
04:1630	869:	1060:	751	409:1	69:0
04:1730	869:	1059:	751	408:1	67:0
04:1830	871:	1060:	751	410:0	67:0
04:1930	865:	1060:	751	407:0	70:0
04:2030	870:	1061:	751	407:0	65:0
04:2130	870:	1060:	771	405:3	64:0
04:2230	875:	1060:	781	405:3	70:0
04:2330	863:	1060:	781	405:3	65:0
05:0030	870:	1060:	781	407:0	70:0
05:0130	870:	1060:	751	405:3	54:0
05:0230	874:	1060:	781	405:3	70:0
05:0330	868:	1060:	781	405:3	70:0
05:0430	868:	1060:	781	407:0	60:0
05:0530	868:	1060:	781	405:5	65:0
05:0630	868:	1060:	781	405:5	69:0
05:0730	868:	1060:	781	405:3	69:0
05:0830	870:	1060:	781	405:1	69:0

Table 6e1

Run 1 Temperatures, Fuel and Blend Rates

DAY:TIME	TEMPERATURE: DEG C:			RATE: LB/HR	
	BASEFEED	BLEND	RECYCLE	OIL	STONE
25:0930	880:	1860:	75:	805:6	65:0
25:1030	883:	1861:	75:	805:6	69:0
25:1130	890:	1860:	10:	805:6	59:0
25:1230	870:	1860:	73:	803:6	67:0
25:1330	870:	1859:	75:	805:6	65:0
25:1430	876:	1860:	75:	805:6	69:0
25:1530	873:	1858:	75:	805:6	77:0
25:1630	878:	1858:	75:	809:8	83:0
25:1730	871:	1863:	75:	810:7	90:0
25:1830	881:	1861:	75:	809:1	93:0
25:1930	879:	1861:	78:	810:9	90:0
25:2030	883:	1860:	78:	810:0	86:0
25:2130	880:	1860:	70:	811:0	91:0
25:2230	881:	1860:	72:	810:9	96:0
25:2330	878:	1858:	70:	809:1	96:0
26:0030	871:	1860:	70:	809:1	98:0
26:0130	870:	1860:	70:	810:0	90:0
26:0230	870:	1860:	70:	810:9	95:0
26:0330	875:	1860:	70:	809:1	93:0
26:0430	880:	1860:	70:	810:0	78:0
26:0530	885:	1860:	70:	810:0	93:0
26:0630	881:	1860:	70:	810:0	97:0
26:0730	886:	1860:	70:	810:0	85:0
26:0830	890:	1860:	70:	810:0	88:0
26:0930	890:	1860:	70:	809:1	90:0
26:1030	883:	1860:	70:	811:0	83:0
26:1130	880:	1860:	72:	810:0	88:0
26:1230	886:	1860:	70:	810:9	85:0
26:1330	889:	1860:	74:	809:1	87:0
26:1430	885:	1860:	74:	810:0	85:0
26:1530	885:	1860:	73:	809:1	90:0
26:1630	885:	1860:	74:	810:0	83:0
26:1730	887:	1860:	74:	809:1	89:0
26:1830	890:	1859:	70:	808:1	87:0

Table D-C

Run 2 Gas Rates

DAY:TIME	GAS RATES: SEEN					SEEN
	DASIFIER		PILOT	REGENERATOR		VELOCITY
	AIR	FLUF GAS	PROPANE	AIR	NITROGEN	FT/SEC
3:0130	878:	89:	2:5	18:6	0:1	8:69
3:0230	878:	89:	2:5	17:5	0:1	8:96
3:0330	887:	117:	2:4	16:7	0:1	1:61
3:0430	858:	111:	2:4	15:8	0:1	8:80
3:0530	858:	111:	2:5	17:5	:1	8:78
3:0630	858:	106:	2:5	15:5	0:1	1:93
3:0730	865:	105:	2:5	14:8	0:1	1:59
3:0930	861:	93:	2:5	15:5	0:1	1:95
3:1030	868:	103:	2:5	17:3	0:1	8:77
3:1130	869:	085:	2:5	16:7	0:1	8:58
3:1230	888:	135:	2:5	19:9	0:1	8:81
3:1330	889:	109:	2:5	16:3	0:1	8:57
3:1430	878:	129:	2:5	18:7	0:1	8:67
3:1530	888:	122:	2:5	18:3	0:1	8:67
3:1630	888:	108:	2:5	18:8	0:1	8:58
3:1730	879:	116:	2:5	15:9	0:1	8:56
3:1830	883:	105:	2:5	16:0	0:1	8:70
3:1930	893:	105:	2:5	15:0	0:1	8:60
3:0030	873:	78:	2:5	17:1	0:1	8:53
3:0130	736:	115:	2:5	13:9	0:1	1:50
3:0230	703:	98:	2:5	18:5	0:1	8:61
3:0330	818:	97:	2:5	08:1	0:1	8:90
3:0430	818:	95:	2:5	15:9	0:1	8:70
3:0530	818:	90:	2:5	15:8	0:1	
3:0630	818:	98:	2:5	19:5	0:1	8:91
3:0730	816:	93:	2:5	19:0	0:1	8:97
3:0830	814:	88:	2:5	08:2	0:1	8:14
3:0930	818:	854:	2:5	08:6	0:1	1:13
3:0A30	819:	80:	2:5	08:6	0:1	8:18
3:0B30	819:	85:	2:5	08:6	0:1	8:06
3:0C30	819:	80:	2:5	08:3	0:1	8:02
3:0D30	868:	81:	2:5	17:5	0:1	8:58
3:0E30	818:	85:	2:5	19:8	0:1	8:98
3:0F30	865:	106:	2:5	19:8	0:1	8:78
3:1030	868:	100:	2:5	19:7	0:1	8:87
3:1130	818:	103:	2:5	08:5	0:1	8:10
3:1230	818:	112:	2:5	18:8	0:1	8:11
3:1330	818:	110:	2:5	18:3	0:1	8:13
3:1430	818:	108:	2:4	17:7	0:1	8:70
3:1730	818:	110:	2:1	15:8	0:1	1:73
3:1830	818:	101:	2:3	19:1	0:1	8:79
3:1930	818:	104:	2:5	17:6	0:1	8:88
3:2030	818:	107:	2:5	18:9	0:1	8:89
3:2130	847:	106:	2:5	18:0	0:1	1:65
3:2230	819:	108:	2:5	18:7	0:1	8:71
3:2330	815:	105:	2:5	08:1	0:1	8:83

Table 10-1

Run 1 Gas Ratios

DAY:TIME	GASIFIER AIR	GASIFIER FLOW GAS	GAS RATES: PILOT PROPANE	SEED REGENERATOR AIR	SEED REGENERATOR SETPOINT%	RECEN VELOCITY FT/SEC
4:00:30	935:	105:	0:5	19:1	0:1	0:70
4:01:30	933:	98:	0:5	19:8	0:1	0:71
4:02:30	936:	99:	0:5	19:7	0:1	0:71
4:03:30	936:	97:	0:5	20:0	0:1	0:70
4:04:30	936:	97:	0:5	20:0	0:1	0:70
4:05:30	939:	110:	0:5	20:0	0:1	0:71
4:06:30	935:	96:	0:5	20:0	0:1	0:71
4:07:30	935:	95:	0:5	20:0	0:1	0:70
4:08:30	935:	100:	0:5	20:0	0:1	0:70
4:09:30	936:	98:	0:5	20:0	0:1	0:69
4:10:30	933:	99:	0:5	19:9	0:1	0:69
4:11:30	933:	110:	0:5	19:9	0:1	0:70
4:12:30	933:	110:	0:5	19:9	0:1	0:70
4:13:30	933:	101:	0:5	19:9	0:1	0:69
4:14:30	933:	111:	0:5	19:9	0:1	0:70
4:15:30	935:	111:	0:5	19:9	0:1	0:70
4:16:30	933:	111:	0:5	19:9	0:1	0:70
4:17:30	933:	111:	0:5	19:9	0:1	0:70
4:18:30	933:	111:	0:5	19:9	0:1	0:70
4:19:30	933:	111:	0:5	19:9	0:1	0:70
4:20:30	933:	111:	0:5	19:9	0:1	0:70
4:21:30	933:	111:	0:5	19:9	0:1	0:70
4:22:30	933:	111:	0:5	19:9	0:1	0:70
4:23:30	933:	111:	0:5	19:9	0:1	0:70
5:00:30	933:	111:	0:5	19:9	0:1	0:70
5:01:30	933:	111:	0:5	19:9	1:0	0:70
5:02:30	933:	111:	0:5	19:9	0:1	0:70
5:03:30	933:	111:	0:5	19:9	1:0	0:70
5:04:30	933:	111:	0:5	19:9	0:1	0:70
5:05:30	933:	110:	0:5	19:9	1:0	0:70
5:06:30	933:	111:	0:5	19:9	0:0	0:70
5:07:30	933:	111:	0:5	19:9	1:0	0:70
5:08:30	933:	111:	0:5	19:9	1:0	0:70
5:09:30	933:	111:	0:5	19:9	1:0	0:70
5:10:30	933:	111:	0:5	19:9	0:0	0:70
5:11:30	933:	111:	0:5	19:9	1:0	0:70
5:12:30	933:	99:	0:5	19:9	1:0	0:70
5:13:30	933:	99:	0:5	19:9	1:0	0:70
5:14:30	933:	99:	0:5	19:9	1:0	0:70
5:15:30	933:	99:	0:5	19:9	1:0	0:70
5:16:30	933:	99:	0:5	19:9	0:0	0:70
6:00:30	933:	100:	0:5	19:9	1:0	0:70
6:01:30	933:	100:	0:5	19:9	0:1	0:70
6:02:30	933:	110:	0:5	19:9	0:1	0:70
6:03:30	933:	110:	0:5	19:9	1:0	0:70
6:04:30	933:	105:	0:5	19:9	1:0	0:70
6:05:30	933:	99:	0:5	19:9	0:0	0:70
6:06:30	933:	99:	0:5	19:9	1:0	0:70

Table D41

Run 5 Gas Rates

DAY:TIME	GAS RATES: CASIFIER AIR	CASIFIER FLUE GAS	PILOT PROPANE	SEPP RECIRCULATION AIR	SEPP RECIRCULATION NITROGEN	GEYES VELOCITY RIZER
9:0930	0.99:	113:	0.5	00.0	1.0	5.09
9:0330	0.98:	113:	0.5	00.1	0.1	5.10
9:0830	0.65:	106:	0.5	01.0	0.1	5.04
9:0130	0.98:	101:	0.5	01.0	1.0	5.05
9:0930	0.99:	100:	0.5	00.5	0.1	5.04
9:0330	0.66:	100:	0.5	00.5	0.1	5.01
9:0830	0.65:	100:	0.5	00.7	0.1	5.01
9:0530	0.66:	98:	0.5	01.1	0.1	5.07
11:1430	0.96:	113:	0.7	01.1	0.1	5.11
11:1530	0.96:	100:	0.7	00.1	0.1	5.03
11:1630	0.99:	99:	0.5	00.0	1.1	5.00
11:1730	0.96:	99:	0.5	00.0	0.1	5.01
11:1830	0.96:	100:	0.5	00.0	0.1	5.01
11:1930	0.69:	98:	0.5	00.0	0.1	5.00
11:2030	0.69:	98:	0.5	00.0	0.1	5.00
11:2130	0.69:	96:	0.5	01.1	0.1	5.01
11:2230	0.99:	98:	0.5	01.0	0.1	5.01
11:2330	0.71:	99:	0.5	10.0	0.1	5.11
12:0030	0.68:	99:	0.5	01.0	0.1	5.01
12:0130	0.68:	97:	0.5	00.0	0.1	5.05
12:0230	0.66:	97:	0.5	10.1	0.1	5.10
12:0330	0.66:	97:	0.5	01.0	0.1	5.11
12:0430	0.66:	97:	0.5	01.1	0.1	5.01
12:0530	0.66:	97:	0.5	01.0	0.1	5.01
12:0630	0.96:	105:	0.5	01.0	0.1	5.11
12:0730	0.91:	111:	0.5	01.5	0.1	5.15
12:0830	0.91:	111:	0.5	01.1	0.1	5.10
12:0930	0.96:	98:	0.5	01.0	0.1	5.10
12:1030	0.96:	98:	0.5	00.1	0.1	5.07
12:1130	0.96:	97:	0.5	00.1	0.1	5.01
12:1230	0.96:	97:	0.5	00.0	0.1	5.00
12:1330	0.91:	99:	0.5	00.0	0.1	5.00
12:1430	0.91:	99:	0.5	00.0	0.1	5.00
12:1530	0.91:	97:	0.5	00.0	0.1	5.00
12:1630	0.91:	97:	0.5	00.0	0.1	5.00
12:1730	0.91:	97:	0.5	00.0	0.1	5.00
12:1830	0.91:	97:	0.5	00.0	0.1	5.00
12:1930	0.91:	97:	0.5	00.0	0.1	5.00
12:2030	0.91:	97:	0.5	00.0	0.1	5.00
12:2130	0.91:	97:	0.5	00.0	0.1	5.00
12:2230	0.91:	97:	0.5	00.0	0.1	5.00
12:2330	0.91:	97:	0.5	00.0	0.1	5.00
13:0030	0.91:	97:	0.5	00.0	0.1	5.00
13:0130	0.91:	97:	0.5	00.0	0.1	5.00
13:0230	0.91:	97:	0.5	00.0	0.1	5.00
13:0330	0.91:	97:	0.5	00.0	0.1	5.00
13:0430	0.91:	97:	0.5	00.0	0.1	5.00
13:0530	0.91:	97:	0.5	00.0	0.1	5.00

Table 3e1

Run 1 Gas Ratio

DAY:TIME	CASIFIER		GAS RATIO: FROM FILE		DILUTION CALCULATED FROM	
	AIR	FLUE GAS	PROPANE	AIR	NITROGEN	TRACER
15:0730	0.61	1.00	0.16	19.15	0.11	0.09
15:0830	0.58	1.03	0.16	19.15	0.11	0.09
15:0930	0.59	1.01	0.16	19.19	0.11	0.09
15:1030	0.58	1.01	0.16	19.19	0.11	0.09
15:1130	0.56	1.07	0.16	19.19	0.11	0.09
15:1230	0.60	1.11	0.16	19.10	0.11	0.09
15:1330	0.69	1.02	0.15	19.11	0.11	0.09
15:1530	0.69	1.00	0.15	19.11	0.11	0.09
15:1630	0.61	0.95	0.15	19.15	0.11	0.09
15:1730	0.60	0.98	0.15	19.15	0.11	0.09
15:1830	0.60	0.98	0.15	19.15	0.11	0.09
15:1930	0.61	0.98	0.15	19.13	0.11	0.09
16:0830	0.56	0.98	0.15	19.18	0.11	0.09
16:2130	0.61	0.99	0.15	19.15	0.11	0.09
16:2330	0.60	0.96	0.15	19.15	0.11	0.09
17:0030	0.60	0.98	0.15	19.15	0.11	0.09
17:0130	0.59	0.98	0.15	19.15	0.11	0.09
17:0230	0.60	0.97	0.15	19.15	0.11	0.09
17:0330	0.60	0.97	0.15	19.15	0.11	0.09
17:0430	0.60	0.97	0.15	19.15	0.11	0.09
17:0530	0.60	0.97	0.15	19.15	0.11	0.09
17:0630	0.60	0.97	0.15	19.15	0.11	0.09
17:0730	0.60	0.97	0.15	19.15	0.11	0.09
17:0830	0.60	0.97	0.15	19.15	0.11	0.09
17:0930	0.60	0.97	0.15	19.15	0.11	0.09
17:1030	0.60	0.97	0.15	19.15	0.11	0.09
17:1130	0.60	0.97	0.15	19.15	0.11	0.09
17:1230	0.60	0.97	0.15	19.15	0.11	0.09
17:1330	0.60	0.97	0.15	19.15	0.11	0.09
17:1430	0.60	0.97	0.15	19.15	0.11	0.09
17:1530	0.60	0.97	0.15	19.15	0.11	0.09
17:1630	0.60	0.97	0.15	19.15	0.11	0.09
17:1730	0.60	0.97	0.15	19.15	0.11	0.09
17:1830	0.60	0.97	0.15	19.15	0.11	0.09
17:1930	0.60	0.97	0.15	19.15	0.11	0.09
17:2030	0.60	0.97	0.15	19.15	0.11	0.09
17:2130	0.60	0.97	0.15	19.15	0.11	0.09
17:2230	0.60	0.97	0.15	19.15	0.11	0.09
17:2330	0.60	0.97	0.15	19.15	0.11	0.09
18:0030	0.60	0.97	0.15	19.15	0.11	0.09
18:0130	0.60	0.97	0.15	19.15	0.11	0.09
18:0230	0.60	0.97	0.15	19.15	0.11	0.09
18:0330	0.60	0.97	0.15	19.15	0.11	0.09
18:0430	0.60	0.97	0.15	19.15	0.11	0.09
18:0530	0.60	0.97	0.15	19.15	0.11	0.09
18:0630	0.60	0.97	0.15	19.15	0.11	0.09
18:0730	0.60	0.97	0.15	19.15	0.11	0.09
18:0830	0.60	0.97	0.15	19.15	0.11	0.09
18:0930	0.60	0.97	0.15	19.15	0.11	0.09
18:1030	0.60	0.97	0.15	19.15	0.11	0.09
18:1130	0.60	0.97	0.15	19.15	0.11	0.09
18:1230	0.60	0.97	0.15	19.15	0.11	0.09
18:1330	0.60	0.97	0.15	19.15	0.11	0.09
18:1430	0.60	0.97	0.15	19.15	0.11	0.09
18:1530	0.60	0.97	0.15	19.15	0.11	0.09
18:1630	0.60	0.97	0.15	19.15	0.11	0.09
18:1730	0.60	0.97	0.15	19.15	0.11	0.09
18:1830	0.60	0.97	0.15	19.15	0.11	0.09
18:1930	0.60	0.97	0.15	19.15	0.11	0.09
18:2030	0.60	0.97	0.15	19.15	0.11	0.09
18:2130	0.60	0.97	0.15	19.15	0.11	0.09
18:2230	0.60	0.97	0.15	19.15	0.11	0.09
18:2330	0.60	0.97	0.15	19.15	0.11	0.09

Table B-1

Run 1 Gas Rates

DAY:TIME	GASEIFIER		GAS RATES: PILOT	SCFH REFRIGERATOR		SECOND VELOCITY
	AIR	FLUE GAS	PROPANE	AIR	SECONDS	FT/SEC
08:2330	883	63	0.8	93.8	120	0.78
01:0030	886	60	0.8	88.5	120	0.73
01:0130	886	60	0.8	88.3	120	0.73
01:0230	893	60	0.8	91.3	120	0.75
01:0330	883	66	0.7	88.8	120	0.73
01:0430	889	66	0.8	91.5	120	0.75
01:0530	885	66	0.8	91.5	120	0.75
01:0630	883	66	0.8	91.3	120	0.75
01:0730	877	66	0.8	90.1	120	0.74
01:0830	880	66	0.8	90.8	120	0.75
01:0930	877	66	0.8	90.3	120	0.74
01:1030	879	66	0.8	90.5	120	0.74
01:1130	880	66	0.8	90.7	120	0.74
01:1230	877	66	0.8	90.1	120	0.74
01:1330	880	66	0.8	90.8	120	0.75
01:1430	869	66	0.8	89.0	120	0.73
01:1530	869	66	0.8	89.0	120	0.73
01:1630	869	66	0.8	89.0	120	0.73
01:1730	869	66	0.8	89.0	120	0.73
01:1830	854	66	0.8	87.5	120	0.72
01:1930	860	66	0.8	88.1	120	0.72
01:2030	856	66	0.8	87.7	120	0.72
01:2130	866	66	0.8	88.7	120	0.73
01:2230	858	66	0.8	87.9	120	0.72
01:2330	858	66	0.8	87.9	120	0.72
02:0030	855	66	0.8	87.6	120	0.72
02:0130	858	66	0.8	87.9	120	0.72
02:0230	858	66	0.8	87.9	120	0.72
02:0330	858	66	0.8	87.9	120	0.72
02:0430	858	66	0.8	87.9	120	0.72
02:0530	858	66	0.8	87.9	120	0.72
02:0630	858	66	0.8	87.9	120	0.72
02:0730	858	66	0.8	87.9	120	0.72
02:0830	858	66	0.8	87.9	120	0.72
02:0930	858	66	0.8	87.9	120	0.72
02:1030	858	66	0.8	87.9	120	0.72
02:1130	858	66	0.8	87.9	120	0.72
02:1230	858	66	0.8	87.9	120	0.72
02:1330	858	66	0.8	87.9	120	0.72
02:1430	858	66	0.8	87.9	120	0.72
02:1530	858	66	0.8	87.9	120	0.72
02:1630	858	66	0.8	87.9	120	0.72
02:1730	858	66	0.8	87.9	120	0.72
02:1830	858	66	0.8	87.9	120	0.72

Table Bed

Bed 1 Gas Rates

DATE/TIME	GAS RATES, SCFH					
	DAESIFIER		PILOT	DETERMINATION		DECEN VELOCITY
	AIR	FLUE GAS	PROPANE	AIR	NITROGEN	FT/SEC
01:17:30	2031	1001	0.7	2113	110	0117
01:18:30	2161	1001	017	2115	110	5100
01:19:30	2131	1001	017	2115	110	5100
01:20:30	2081	1001	110	2115	110	0117
01:21:30	2031	1031	016	2015	110	0121
01:22:30	2131	1051	016	2016	110	0163
01:23:30	2181	1051	016	2018	110	0110
01:24:30	2071	1031	016	2013	110	0170
01:25:30	2101	1031	016	2019	110	0109
01:26:30	2101	971	016	2015	110	0100
01:27:30	2101	971	016	2011	110	0171
01:28:30	2101	971	016	2016	110	0177
01:29:30	2151	971	016	2019	110	0167
01:30:30	2191	971	016	2021	113	0117
01:31:30	2151	901	015	2017	113	0173
01:32:30	2141	951	015	2010	113	0117
01:33:30	2081	951	015	2013	113	0111
01:34:30	2134	951	015	2015	113	0161
01:35:30	2101	971	017	2110	113	0101
01:36:30	2051	971	017	2110	113	0117
01:37:30	2051	971	017	211	113	0117
01:38:30	2051	971	017	2110	113	0100
01:39:30	2051	971	017	2110	113	0100
01:40:30	2051	971	017	2110	113	0100
01:41:30	2051	971	017	2110	113	0100
01:42:30	2051	971	017	2110	113	0100
01:43:30	2051	971	017	2110	113	0100
01:44:30	2051	971	017	2110	113	0100
01:45:30	2051	971	017	2110	113	0100
01:46:30	2051	971	017	2110	113	0100
01:47:30	2051	971	017	2110	113	0100
01:48:30	2051	971	017	2110	113	0100
01:49:30	2051	971	017	2110	113	0100
01:50:30	2051	971	017	2110	113	0100
01:51:30	2051	971	017	2110	113	0100
01:52:30	2051	971	017	2110	113	0100
01:53:30	2051	971	017	2110	113	0100
01:54:30	2051	971	017	2110	113	0100
01:55:30	2051	971	017	2110	113	0100
01:56:30	2051	971	017	2110	113	0100
01:57:30	2051	971	017	2110	113	0100
01:58:30	2051	971	017	2110	113	0100
01:59:30	2051	971	017	2110	113	0100
02:00:30	2051	971	017	2110	113	0100
02:01:30	2051	971	017	2110	113	0100
02:02:30	2051	971	017	2110	113	0100

Table D-1

Run 1 Gas Rates

DAY:TIME	CASIFIER AIR	CASIFIER FLUE GAS	PILOT PROPANE	GEN DEGENERATOR AIR	GEN DEGENERATOR NITROGEN	RE-ED VELOCITY FT/SEC
25:0930	3151	911	045	0145	140	5.01
25:1030	0051	911	045	0143	140	5.10
25:1130	0151	901	045	0143	140	5.105
25:1230	0134	911	045	0143	140	5.105
25:1330	0054	911	045	0140	140	5.133
25:1430	0061	911	045	0140	140	5.133
25:1530	0061	911	045	0140	140	5.133
25:1630	0051	911	045	0140	140	5.133
25:1730	0051	911	045	0140	140	5.101
25:1830	0061	911	045	0140	140	5.101
25:1930	0051	911	045	0140	140	5.105
25:0030	0051	911	045	0140	140	5.105
25:0130	0111	941	045	0140	140	5.101
25:0230	0111	941	045	0140	140	5.101
25:0330	0151	911	045	0140	140	5.101
26:0030	0101	911	045	0140	140	5.101
26:0130	0051	911	045	0141	140	5.101
26:0230	0061	911	045	0141	140	5.101
26:0330	0061	911	045	0141	140	5.101
26:0430	0051	911	045	0141	140	5.101
26:0530	0051	911	045	0141	140	5.101
26:0630	0061	911	045	0141	140	5.101
26:0730	0061	911	045	0141	140	5.101
26:0830	0051	911	045	0141	140	5.101
26:0930	0051	911	045	0141	140	5.101
26:1030	0121	911	045	0141	140	5.101
26:1130	0061	911	045	0141	140	5.101
26:1230	0161	951	045	0141	140	5.101
26:1330	0161	911	045	0141	140	5.101
26:1430	0051	901	045	0140	140	5.101
26:1530	0051	911	045	0140	140	5.101
26:1630	0151	911	045	0140	140	5.101
26:1730	0051	911	045	0140	140	5.101
26:1830	0051	911	045	0140	140	5.101

Table [illegible]

Run 1 /AFB Pressures

DAY:TIME	CAVITIES PRESSURES (PSI GIG)				
	GAS SPACE	DISTRIB	RED D(P)	SEC GAS	RED D(P)
5:01:30	14.0	14.0	15.0	0.10	15.0
5:03:30	15.0	14.0	20.0	0.10	21.0
5:04:30	16.0	15.0	19.0	0.10	25.0
5:06:30	17.0	15.0	19.5	0.10	27.0
5:07:30	17.5	15.0	19.5	0.10	29.0
5:08:30	17.5	16.0	19.5	0.15	30.0
5:09:30	17.5	16.0	19.0	0.15	31.0
5:10:30	18.0	16.0	19.5	0.15	32.0
5:11:30	18.5	16.0	19.5	0.20	33.0
5:12:30	18.0	16.0	19.0	0.20	34.0
5:13:30	18.0	16.0	19.0	0.20	35.0
5:14:30	18.5	16.0	19.5	0.20	36.0
5:15:30	18.5	16.5	19.0	0.20	37.0
5:16:30	18.5	16.5	19.0	0.20	38.0
5:17:30	18.5	16.5	19.5	0.20	39.0
5:18:30	18.0	16.5	19.5	0.20	40.0
5:19:30	18.0	16.5	19.5	0.20	41.0
5:20:30	18.5	16.5	19.5	0.20	42.0
5:21:30	18.5	16.5	19.5	0.20	43.0
5:22:30	18.5	16.5	19.5	0.20	44.0
5:23:30	18.5	16.5	19.5	0.20	45.0
5:24:30	18.5	16.5	19.5	0.20	46.0
5:25:30	18.5	16.5	19.5	0.20	47.0
5:26:30	18.5	16.5	19.5	0.20	48.0
5:27:30	18.5	16.5	19.5	0.20	49.0
5:28:30	18.5	16.5	19.5	0.20	50.0
5:29:30	18.5	16.5	19.5	0.20	51.0
5:30:30	18.5	16.5	19.5	0.20	52.0
5:31:30	18.5	16.5	19.5	0.20	53.0
5:32:30	18.5	16.5	19.5	0.20	54.0
5:33:30	18.5	16.5	19.5	0.20	55.0
5:34:30	18.5	16.5	19.5	0.20	56.0
5:35:30	18.5	16.5	19.5	0.20	57.0
5:36:30	18.5	16.5	19.5	0.20	58.0
5:37:30	18.5	16.5	19.5	0.20	59.0
5:38:30	18.5	16.5	19.5	0.20	60.0
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10:38:30	18.5	16.5	19.5	0.20	360.0
10:39:30	18.5	16.5	19.5	0.20	361.0
10:40:30	18.5	16.5	19.5	0.20	362.0
10:41:30	18.5	16.5	19.5	0.20	363.0
10:42:30	18.5	16.5	19.5	0.20	364.0
10:43:30	18.5	16.5	19.5	0.20	365.0
10:44:30	18.5	16.5	19.5	0.20	366.0
10:45:30	18.5	16.5	19.5	0.20	367.0
10:46:30	18.5	16.5	19.5	0.20	368.0
10:47:30	18.5	16.5	19.5	0.20	369.0
10:48:30	18.5	16.5	19.5	0.20	370.0
10:49:30	18.5	16.5	19.5	0.20	371.0
10:50:30	18.5	16.5	19.5	0.20	372.0
10:51:30	18.5	16.5	19.5	0.20	373.0
10:52:30	18.5	16.5	19.5	0.20	374.0
10:53:30	18.5	16.5	19.5	0.20	375.0
10:54:30	18.5	16.5	19.5	0.20	376.0
10:55:30	18.5	16.5	19.5	0.20	377.0
10:56:30	18.5	16.5	19.5	0.20	378.0
10:57:30	18.5	16.5	19.5	0.20	379.0
10:58:30	18.5	16.5	19.5	0.20	380.0
10:59:30	18.5	16.5	19.5	0.20	381.0
11:00:30	18.5	16.5	19.5	0.20	382.0
11:01:30	18.5	16.5	19.5	0.20	383.0
11:02:30	18.5	16.5	19.5	0.20	384.0
11:03:30	18.5	16.5	19.5	0.20	385.0
11:04:30	18.5	16.5	19.5	0.20	386.0
11:05:30	18.5	16.5	19.5	0.20	387.0
11:06:30	18.5	16.5	19.5	0.20	388.0
11:07:30	18.5	16.5	19.5	0.20	389.0
11:08:30	18.5	16.5	19.5	0.20	390.0
11:09:30	18.5	16.5	19.5	0.20	391.0
11:10:30	18.5	16.5	19.5	0.20	392.0
11:11:30	18.5	16.5	19.5	0.20	393.0
11:12:30	18.5	16.5	19.5	0.20	394.0
11:13:30	18.5	16.5	19.5	0.20	395.0
11:14:30	18.5	16.5	19.5	0.20	396.0
11:15:30	18.5	16.5	19.5	0.20	397.0
11:16:30	18.5	16.5	19.5	0.20	398.0
11:17:30	18.5	16.5	19.5	0.20	399.0
11:18:30	18.5	16.5	19.5	0.20	400.0
11:19:30	18.5	16.5	19.5	0.20	401.0
11:20:30	18.5	16.5	19.5	0.20	402.0
11:21:30	18.5	16.5	19.5	0.20	403.0
11:22:30	18.5	16.5	19.5	0.20	404.0
11:23:30	18.5	16.5	19.5	0.20	405.0
11:24:30	18.5	16.5	19.5	0.20	406.0
11:25:30	18.5	16.5	19.5	0.20	407.0
11:26:30	18.5	16.5	19.5	0.20	408.0
11:27:30	18.5	16.5	19.5	0.20	409.0
11:28:30	18.5	16.5	19.5	0.20	410.0
11:29:30	18.5	16.5	19.5	0.20	411.0
11:30:30	18.5	16.5	19.5	0.20	412.0
11:31:30	18.5	16.5	19.5	0.20	413.0
11:32:30	18.5	16.5	19.5	0.20	414.0
11:33:30	18.5	16.5	19.5	0.20	415.0
11:34:30	18.5	16.5	19.5	0.20	416.0
11:35:30	18.5	16.5	19.5	0.20	417.0
11:36:30	18.5	16.5	19.5	0.20	418.0
11:37:30	18.5	16.5	19.5	0.20	419.0
11:38:30	18.5	16.5	19.5	0.20	420.0
11:39:30	18.5	16.5	19.5	0.20	421.0

Table No.1

Run 1 CAPD Pressures

DAY.TIME	BASIFLEX PRESSURES (PSI G/CM)				
	BAG SPACE	DIALYSIS	BED	DP1	DP2
		D1	D2		
1.00:30	18.0	13.5	18.0	0.50	0.0
1.01:30	16.3	13.0	18.0	0.50	0.0
1.02:30	16.0	13.0	18.0	0.50	0.0
1.03:30	17.1	13.5	17.5	0.50	0.5
1.04:30	18.0	13.5	17.5	0.50	0.5
1.05:30	18.0	13.5	17.0	0.75	0.0
1.06:30	19.0	13.5	17.0	0.75	0.0
1.07:30	19.0	13.5	19.0	0.75	0.0
1.08:30	19.5	13.5	19.3	0.75	0.0
1.09:30	20.0	15.0	18.0	0.50	0.0
1.10:30	20.0	15.0	18.0	0.50	0.0
1.11:30	20.0	15.0	18.0	0.50	0.0
1.12:30	20.5	15.0	17.5	0.50	0.5
1.13:30	20.5	16.0	18.0	0.50	0.0
1.14:30	21.0	16.0	18.0	0.50	0.0
1.15:30	20.5	16.0	18.0	0.50	0.0
1.16:30	21.0	16.0	18.0	0.50	0.0
1.17:30	21.0	16.0	18.0	0.50	0.0
1.18:30	21.0	16.0	18.0	0.50	0.0
1.19:30	21.0	17.0	18.0	0.50	0.0
1.20:30	21.0	17.0	18.0	0.50	0.0
1.21:30	21.5	17.0	18.0	0.50	0.0
1.22:30	20.5	17.0	18.0	0.50	0.0
1.23:30	20.5	17.5	18.0	0.25	0.0
2.00:30	20.5	15.0	18.0	0.25	0.0
2.01:30	20.5	15.5	18.0	0.25	0.0
2.02:30	20.5	15.5	18.0	0.25	0.0
2.03:30	20.0	15.0	18.0	0.25	0.0
2.04:30	20.5	15.0	18.0	0.25	0.0
2.05:30	20.7	15.0	18.0	0.25	0.0
2.06:30	20.3	15.0	17.7	0.25	0.0
2.07:30	21.5	15.5	18.0	0.25	0.0
2.08:30	21.0	16.5	18.0	0.50	0.0
2.09:30	20.5	16.0	18.0	0.50	0.0
2.10:30	20.5	16.0	18.0	0.50	0.0
2.11:30	20.0	15.0	18.0	0.50	0.0
2.12:30	19.5	15.0	18.0	0.50	0.0
2.13:30	20.5	15.5	17.5	0.25	0.0
2.14:30	20.5	15.5	17.5	0.25	0.0
2.15:30	21.0	15.5	17.5	0.25	0.0
2.16:30	20.5	15.5	17.5	0.25	0.0
2.17:30	20.0	15.0	17.5	0.25	0.0
2.18:30	19.5	15.5	17.0	0.25	0.0
2.19:30	20.0	15.5	17.5	0.25	0.0
2.20:30	20.0	15.5	18.0	0.25	0.0
2.21:30	20.0	15.0	18.0	0.50	0.0
2.22:30	20.0	15.0	18.0	0.50	0.0
2.23:30	21.0	15.5	18.0	0.25	0.0

Table D-1

Vol 1 (APR Pressure)

DAY:TIME	CASIFIER PRESSURES (PSI GAGE)					DELTA
	GAS SPACE	DIFFUSER D:P:	RED D:P:	DP: (PSI)		RED D:P:
8:00:00	13:0	10:0	10:5	0:5		
8:30:00	13:0	11:0	10:5	0:5		
9:00:00	13:0	12:0	10:0	0:5		
9:01:00	13:0	12:0	10:5	0:5		
9:05:00	13:0	12:0	09:0	0:5		
9:09:00	13:0	12:0	08:0	0:5		
9:09:30	13:0	12:0	08:0	0:5		
9:09:50	13:0	12:0	08:0	0:5		
11:10:00	15:5		10:0	0:5		12:5
11:15:00	15:5		10:5	0:5		13:0
11:16:00	15:0		13:5	0:5		13:5
11:17:00	15:0		13:5	0:5		13:0
11:18:00	15:5		13:5	0:5		13:0
11:19:00	15:0		13:0	0:5		13:0
11:20:00	15:0		12:0	0:5		13:0
11:21:00	15:0		12:5	0:5		13:0
11:22:00	15:0		12:5	0:5		13:0
11:23:00	15:0		12:5	0:5		13:0
12:00:00	15:0	11:0	12:5	0:5		13:0
12:01:00	15:5	11:0	12:0	0:5		13:0
12:10:00	15:5	11:0	12:0	0:5		13:0
12:20:00	15:5	11:0	11:5	0:5		13:0
12:30:00	15:0	11:0	11:5	0:5		13:0
12:40:00	15:0	11:0	11:5	0:5		13:0
12:50:00	15:0	11:0	11:5	0:5		13:0
13:00:00	15:0	11:0	11:5	0:5		13:0
13:10:00	15:0	11:0	11:5	0:5		13:0
13:20:00	15:0	11:0	11:5	0:5		13:0
13:30:00	15:0	11:0	11:5	0:5		13:0
13:40:00	15:0	11:0	11:5	0:5		13:0
13:50:00	15:0	11:0	11:5	0:5		13:0
14:00:00	15:0	11:0	11:5	0:5		13:0
14:10:00	15:0	11:0	11:5	0:5		13:0
14:20:00	15:0	11:0	11:5	0:5		13:0
14:30:00	15:0	11:0	11:5	0:5		13:0
14:40:00	15:0	11:0	11:5	0:5		13:0
14:50:00	15:0	11:0	11:5	0:5		13:0
15:00:00	15:0	11:0	11:5	0:5		13:0
15:10:00	15:0	11:0	11:5	0:5		13:0
15:20:00	15:0	11:0	11:5	0:5		13:0
15:30:00	15:0	11:0	11:5	0:5		13:0
15:40:00	15:0	11:0	11:5	0:5		13:0
15:50:00	15:0	11:0	11:5	0:5		13:0
16:00:00	15:0	11:0	11:5	0:5		13:0
16:10:00	15:0	11:0	11:5	0:5		13:0
16:20:00	15:0	11:0	11:5	0:5		13:0
16:30:00	15:0	11:0	11:5	0:5		13:0
16:40:00	15:0	11:0	11:5	0:5		13:0
16:50:00	15:0	11:0	11:5	0:5		13:0
17:00:00	15:0	11:0	11:5	0:5		13:0
17:10:00	15:0	11:0	11:5	0:5		13:0
17:20:00	15:0	11:0	11:5	0:5		13:0
17:30:00	15:0	11:0	11:5	0:5		13:0
17:40:00	15:0	11:0	11:5	0:5		13:0
17:50:00	15:0	11:0	11:5	0:5		13:0
18:00:00	15:0	11:0	11:5	0:5		13:0
18:10:00	15:0	11:0	11:5	0:5		13:0
18:20:00	15:0	11:0	11:5	0:5		13:0
18:30:00	15:0	11:0	11:5	0:5		13:0
18:40:00	15:0	11:0	11:5	0:5		13:0
18:50:00	15:0	11:0	11:5	0:5		13:0
19:00:00	15:0	11:0	11:5	0:5		13:0
19:10:00	15:0	11:0	11:5	0:5		13:0
19:20:00	15:0	11:0	11:5	0:5		13:0
19:30:00	15:0	11:0	11:5	0:5		13:0
19:40:00	15:0	11:0	11:5	0:5		13:0
19:50:00	15:0	11:0	11:5	0:5		13:0
20:00:00	15:0	11:0	11:5	0:5		13:0
20:10:00	15:0	11:0	11:5	0:5		13:0
20:20:00	15:0	11:0	11:5	0:5		13:0
20:30:00	15:0	11:0	11:5	0:5		13:0
20:40:00	15:0	11:0	11:5	0:5		13:0
20:50:00	15:0	11:0	11:5	0:5		13:0
21:00:00	15:0	11:0	11:5	0:5		13:0
21:10:00	15:0	11:0	11:5	0:5		13:0
21:20:00	15:0	11:0	11:5	0:5		13:0
21:30:00	15:0	11:0	11:5	0:5		13:0
21:40:00	15:0	11:0	11:5	0:5		13:0
21:50:00	15:0	11:0	11:5	0:5		13:0
22:00:00	15:0	11:0	11:5	0:5		13:0
22:10:00	15:0	11:0	11:5	0:5		13:0
22:20:00	15:0	11:0	11:5	0:5		13:0
22:30:00	15:0	11:0	11:5	0:5		13:0
22:40:00	15:0	11:0	11:5	0:5		13:0
22:50:00	15:0	11:0	11:5	0:5		13:0
23:00:00	15:0	11:0	11:5	0:5		13:0
23:10:00	15:0	11:0	11:5	0:5		13:0
23:20:00	15:0	11:0	11:5	0:5		13:0
23:30:00	15:0	11:0	11:5	0:5		13:0
23:40:00	15:0	11:0	11:5	0:5		13:0
23:50:00	15:0	11:0	11:5	0:5		13:0

Table D-1

Run 1 (AFB Pressures

DAY:TIME	DAYTIME PRESSURES: (Hg)				NIGHT
	OAT SPACE	DISTRIB P_1	RED P_2	CP_1 CP_2	RED P_2
14:07:55	17.0	28.5	11.0	0.75	26.0
14:08:10	16.0	17.5	10.0	0.75	10.0
14:09:55	16.0	28.0	19.5	0.75	15.0
14:10:10	16.0	28.5	19.5	0.75	26.0
14:11:10	16.0	29.0	20.0	0.75	16.0
14:12:55	16.0	28.5	21.0	0.75	20.0
14:13:10	16.0	28.0	21.0	0.75	21.0
14:15:55	15.0	28.0	23.5	0.75	23.5
14:16:10	16.0	19.0	21.0	0.75	20.0
14:17:10	16.0	19.0	21.0	0.75	21.0
14:18:55	16.0	19.5	22.0	0.75	21.0
14:19:10	17.0	19.0	24.0	0.75	23.0
14:20:55	17.0	28.0	24.5	0.75	28.5
14:21:10	17.0	28.0	24.0	0.75	27.0
14:22:10	17.0	28.0	25.0	0.75	27.0
14:23:55	17.0	19.0	25.0	0.75	21.0
14:00:10	17.0	28.0	25.0	0.75	28.0
14:01:10	17.5	28.0	25.0	0.75	27.0
14:02:55	17.5	28.0	26.0	0.75	28.0
14:03:10	17.5	19.5	26.0	0.75	27.0
14:04:10	17.5	28.0	26.0	0.75	27.0
14:05:55	17.5	28.0	26.0	0.75	27.0
14:06:10	17.5	19.5	25.5	0.75	23.0
14:07:10	17.5	19.5	25.5	0.75	23.0
14:08:55	17.5	19.5	25.0	0.75	23.0
14:09:10	17.5	28.0	25.0	0.75	28.0
14:10:10	17.5	19.5	25.0	0.75	23.0
14:11:10	17.5	19.5	25.0	0.75	23.0
14:12:55	17.5	19.0	25.0	0.75	23.0
14:13:10	17.5	19.0	25.0	0.75	23.0
14:14:10	17.5	19.0	25.0	0.75	23.0
14:15:55	17.5	19.0	25.0	0.75	23.0
14:16:10	17.5	19.0	25.0	0.75	23.0
14:17:10	17.5	19.0	25.0	0.75	23.0
14:18:55	17.5	19.0	25.0	0.75	23.0
14:19:10	17.5	19.0	25.0	0.75	23.0
14:20:10	17.5	19.0	25.0	0.75	23.0
14:21:55	17.5	19.0	25.0	0.75	23.0
14:22:10	17.5	19.0	25.0	0.75	23.0
14:23:10	17.5	19.0	25.0	0.75	23.0
14:00:55	17.5	19.0	25.0	0.75	23.0
14:01:10	17.5	19.0	25.0	0.75	23.0
14:02:10	17.5	19.0	25.0	0.75	23.0
14:03:55	17.5	19.0	25.0	0.75	23.0
14:04:10	17.5	19.0	25.0	0.75	23.0
14:05:10	17.5	19.0	25.0	0.75	23.0
14:06:55	17.5	19.0	25.0	0.75	23.0
14:07:10	17.5	19.0	25.0	0.75	23.0
14:08:10	17.5	19.0	25.0	0.75	23.0
14:09:55	17.5	19.0	25.0	0.75	23.0
14:10:10	17.5	19.0	25.0	0.75	23.0
14:11:10	17.5	19.0

Table D-7

Run 5 (APS) Pressures

DAY:TIME	CASTILE PRESSURES (PSIG)				
	GAS SPACE	DISTRIB	BED	DP: DB:	BRINE BED
	D:P:		D:P:		D:P:
00:0430	83.0	19.0	98.0	0.75	98.0
01:0030	83.5	18.0	98.5	0.00	98.0
01:0130	84.0	17.5	98.5	0.75	96.0
01:0230	84.0	17.5	97.0	0.50	96.0
01:0330	84.5	16.0	97.0	0.50	97.0
01:0430	84.5	18.0	98.0	0.75	95.0
01:0530	84.5	15.0	98.0	0.00	97.0
01:0630	85.0	15.0	98.0	0.75	94.0
01:0730	85.5	17.5	98.0	0.75	94.0
01:0830	85.0	17.0	98.0	0.00	97.0
01:0930	84.0	17.5	98.0	0.75	97.0
01:1030	84.5	17.0	98.5	0.75	97.0
01:1130	84.5	17.0	98.0	0.00	96.0
01:1230	84.0	16.5	98.5	0.75	95.0
01:1330	85.5	17.5	98.0	0.00	96.0
01:1430	85.5	17.5	98.0	0.00	97.0
01:1530	86.0	16.0	97.0	0.75	93.0
01:1630	86.5	16.5	98.5	0.75	95.0
01:1730	87.0	16.5	98.5	0.75	98.0
01:1830	87.5	17.5	97.0	0.75	98.0
01:1930	88.5	16.0	98.5	0.00	98.0
01:2030	88.5	17.5	97.5	0.75	95.0
01:2130	88.5	18.0	98.0	0.00	94.0
01:2230	88.5	18.0	98.0	0.75	94.0
01:2330	88.5	18.0	97.5	0.00	96.0
02:0030	89.0	18.0	98.0	0.75	97.0
02:0130	89.0	18.0	97.0	0.00	95.0
02:0230	89.5	18.0	98.5	0.75	95.0
02:0330	89.5	18.0	97.0	0.00	96.0
02:0430	89.5	18.0	97.0	0.00	96.0
02:0530	89.5	18.0	97.0	0.00	95.0
02:0630	89.5	18.0	97.0	0.00	95.0
02:0730	89.5	18.0	97.0	0.00	95.0
02:0830	89.5	18.0	97.0	0.00	95.0
02:0930	89.5	18.0	97.0	0.00	95.0
02:1030	89.5	18.0	97.0	0.00	95.0
02:1130	89.5	18.0	97.0	0.00	95.0
02:1230	89.5	18.0	97.0	0.00	95.0
02:1330	89.5	18.0	97.0	0.00	95.0
02:1430	89.5	18.0	97.0	0.00	95.0
02:1530	89.5	18.0	97.0	0.00	95.0
02:1630	89.5	18.0	97.0	0.00	95.0
02:1730	89.5	18.0	97.0	0.00	95.0
02:1830	89.5	18.0	97.0	0.00	95.0
02:1930	89.5	18.0	97.0	0.00	95.0
02:2030	89.5	18.0	97.0	0.00	95.0
02:2130	89.5	18.0	97.0	0.00	95.0
02:2230	89.5	18.0	97.0	0.00	95.0
02:2330	89.5	18.0	97.0	0.00	95.0

Table D-7

Run 3 CAFD Pressures

DAY:TIME	GASIFIER PRESSURES, IN: H:O:					REFER
	GAS SPACE	RESTRIB	BED	SP: CR:		BED
		B:P:	B:P:			B:P:
03:1730	91:0	84:0	19:5	0:10		90:0
03:1830	91:5	84:0	90:0	0:10		89:0
03:1930	91:0	84:0	91:0	0:10		93:0
03:2000	91:0	84:0	91:5	0:10		93:0
03:2130	90:5	83:5	89:0	0:10		89:0
03:2200	91:0	84:0	93:0	0:10		90:0
03:2330	91:5	84:5	93:0	0:10		94:0
04:0030	91:5	85:0	95:0	0:10		95:0
04:0130	92:0	85:0	93:0	0:10		94:0
04:0230	92:0	85:0	94:0	0:10		94:0
04:0330	92:5	85:0	96:0	0:10		96:0
04:0430	92:0	85:0	94:0	0:10		94:0
04:0530	92:0	84:5	94:0	0:10		94:0
04:0630	92:5	84:5	96:0	0:10		96:0
04:0730	93:0	84:5	96:0	0:10		96:0
04:0830	93:0	85:5	96:0	0:10		96:0
04:0930	94:0	85:0	95:5	0:10		96:0
04:1030	94:0	85:5	95:5	0:10		96:0
04:1130	94:0	85:0	95:5	0:10		96:0
04:1230	94:5	85:5	95:0	0:10		96:0
04:1330	94:0	85:5	95:5	0:10		96:0
04:1430	95:5	85:5	95:0	0:10		96:0
04:1530	95:0	85:5	94:5	0:10		96:0
04:1630	95:0	85:0	94:5	0:10		95:0
04:1730	95:5	85:0	94:0	0:10		95:0
04:1830	95:5	85:0	95:0	0:10		96:0
04:1930	95:0	85:0	95:0	0:10		95:0
04:2030	94:0	84:5	95:0	0:10		95:0
04:2130	94:0	84:5	95:0	0:10		95:0
04:2230	94:0	84:5	95:0	0:10		95:0
04:2330	94:5	84:5	95:0	0:10		95:0
05:0030	94:5	84:5	95:0	0:10		95:0
05:0130	94:0	84:5	95:0	0:10		95:0
05:0230	94:5	84:5	95:0	0:10		95:0
05:0330	94:5	84:5	95:5	0:10		96:0
05:0430	94:5	84:5	95:0	0:10		95:0
05:0530	94:5	84:5	95:0	0:10		95:0
05:0630	94:5	84:5	95:0	0:10		95:0
05:0730	94:5	84:5	95:0	0:10		95:0

Table D-3

Run 1 (API Standard)

DAY:TIME	DAYLIFTER PRESSURES (PSIG)			INJ. WELL		RECEIV
	GAS SPACE	CASING	TBG			GEO DIF
05:0930	3410	3710	3915	8:10		0910
05:1030	3410	3710	3910	8:10		0910
05:1130	3415	3710	3810	8:10		0810
05:1230	3415	3710	3810	8:15		0510
05:1330	3415	3710	3915	8:10		0510
05:1430	3410	3710	3915	8:10		0810
05:1530	3410	3710	3915	8:10		0810
05:1630	3410	3710	3915	8:10		0810
05:1730	3410	3710	3915	8:10		0810
05:1830	3410	3710	3915	8:10		0810
05:1930	3410	3710	3915	8:10		0810
05:2030	3410	3710	3915	8:10		0810
05:2130	3410	3710	3915	8:10		0810
05:2230	3410	3710	3915	8:10		0810
05:2330	3410	3710	3915	8:10		0810
06:0030	3410	3710	3915	8:10		0810
06:0130	3410	3710	3915	8:10		0810
06:0230	3410	3710	3915	8:10		0810
06:0330	3410	3710	3915	8:10		0810
06:0430	3410	3710	3915	8:10		0810
06:0530	3410	3710	3915	8:10		0810
06:0630	3410	3710	3915	8:10		0810
06:0730	3410	3710	3915	8:10		0810
06:0830	3410	3710	3915	8:10		0810
06:0930	3410	3710	3915	8:10		0810
06:1030	3410	3710	3915	8:10		0810
06:1130	3410	3710	3915	8:10		0810
06:1230	3410	3710	3915	8:10		0810
06:1330	3410	3710	3915	8:10		0810
06:1430	3410	3710	3915	8:10		0810
06:1530	3410	3710	3915	8:10		0810
06:1630	3410	3710	3915	8:10		0810
06:1730	3410	3710	3915	8:10		0810
06:1830	3410	3710	3915	8:10		0810
06:1930	3410	3710	3915	8:10		0810
06:2030	3410	3710	3915	8:10		0810
06:2130	3410	3710	3915	8:10		0810
06:2230	3410	3710	3915	8:10		0810
06:2330	3410	3710	3915	8:10		0810
07:0030	3410	3710	3915	8:10		0810
07:0130	3410	3710	3915	8:10		0810
07:0230	3410	3710	3915	8:10		0810
07:0330	3410	3710	3915	8:10		0810
07:0430	3410	3710	3915	8:10		0810
07:0530	3410	3710	3915	8:10		0810
07:0630	3410	3710	3915	8:10		0810
07:0730	3410	3710	3915	8:10		0810
07:0830	3410	3710	3915	8:10		0810
07:0930	3410	3710	3915	8:10		0810
07:1030	3410	3710	3915	8:10		0810
07:1130	3410	3710	3915	8:10		0810
07:1230	3410	3710	3915	8:10		0810
07:1330	3410	3710	3915	8:10		0810
07:1430	3410	3710	3915	8:10		0810
07:1530	3410	3710	3915	8:10		0810
07:1630	3410	3710	3915	8:10		0810
07:1730	3410	3710	3915	8:10		0810
07:1830	3410	3710	3915	8:10		0810
07:1930	3410	3710	3915	8:10		0810
07:2030	3410	3710	3915	8:10		0810
07:2130	3410	3710	3915	8:10		0810
07:2230	3410	3710	3915	8:10		0810
07:2330	3410	3710	3915	8:10		0810

Table 3-5

Computed Boiler Circulation

CAFD Run 3

DAY:HOUR	SOLIDS CIRCULATION: LB/HR		
	CIRCULATING END	TOTAL TO REGEN	SOLIDS FROM REGEN
0:01:30	913	936	946
0:03:30	796	801	807
0:04:30	659	670	670
0:06:30	876	883	901
0:05:30	1103	1106	1106
0:06:30	865	865	868
0:07:30	788	760	766
0:09:30	849	863	870
0:10:30	1073	1055	1101
0:11:30	930	953	961
0:13:30	785	708	731
0:14:30	785	708	730
0:15:30	856	789	787
0:16:30	898	788	785
0:17:30	706	789	751
0:18:30	717	761	763
0:19:30	576	596	596
0:20:30	576	596	596
0:21:30	606	647	657
0:22:30	686	686	686
0:23:30	670	696	696
1:00:30	783	769	771
1:01:30	736	750	763
1:02:30	644	670	668
1:03:30	687	705	710
1:04:30	780	723	730
1:05:30	756	780	780
1:06:30	715	681	681
1:07:30	758	780	783
1:08:30	780	686	689
1:09:30	781	786	786
1:10:30	618	635	636
1:11:30	781	686	615
1:12:30	783	723	778
1:13:30	786	783	786
1:14:30	685	689	618
1:15:30	675	698	705
1:16:30	668	686	693
1:17:30	673	690	691

Table D-6

Simulated Solids Circulation

(APS Run 3)

DAY:HOUR:MIN	SOLIDS CIRCULATION: LB/HR		
	CIRCULATING CAO	TOTAL TO BEDEN	SOLIDS FROM BEDEN
1:18:30	676	598	100
1:19:30	638	668	663
1:20:30	793	816	830
1:21:30	521	517	500
1:22:30	686	710	715
1:23:30	818	835	851
1:00:30	813	839	887
1:01:30	618	638	635
1:02:30	568	558	580
1:03:30	717	782	785
1:04:30	583	573	632
1:05:30	630	859	835
1:06:30	611	636	602
1:07:30	556	585	580
1:08:30	578	616	613
1:09:30	492	548	457
1:10:30	600	556	575
1:11:30	555	581	578
1:12:30	747	769	715
1:13:30			
1:14:30	597	650	587
1:15:30	693	691	633
1:16:30	558	581	597
1:17:30	668	684	691
1:18:30	602	618	615
1:19:30	611	613	611
1:20:30	676	630	609
1:21:30	591	606	595
1:22:30	650	648	684
1:23:30	638	654	655
1:00:30	668	661	651
1:01:30			
1:02:30	693	715	730
1:03:30	651	718	771
1:04:30	593	618	598
1:05:30	658	685	684
1:06:30	631	658	638
1:07:30	618	645	660
1:08:30	613	637	630
1:09:30	566	578	581

Table D-8

Computed Bolide Circulation

(AFB Run 1)

SOLIDS CIRCULATIONS: LBS/HR

DAY:HOUR	CIRCULATING CAD	TOTAL TO BEDEN	SOLIDS FROM BEDEN
5:1030	691	697	695
5:1130	578	590	586
5:1230	636	634	634
5:1330	615	660	630
5:1530	638	650	650
5:1630	690	704	706
6:0030	650	668	660
6:0130	549	550	559
6:0630	585	567	575
6:0830	510	539	540
6:0930	632	659	657
6:1030	509	530	530
6:1130	657	685	680
8:2230	1369	1598	1616
8:2330	1109	1135	1155
9:0030	901	930	954
9:0130	1003	1031	1049
9:0230	1049	1079	1093
9:0330	1007	1056	1080
9:0630	997	1030	1040
9:0530	1050	1084	1093
11:1430	696	700	733
11:1530	700	740	765
11:1630	700	729	736
11:1730	755	783	797
11:1830	683	713	715
11:1930	715	754	761
11:2030	678	700	700
11:2130	796	805	831
11:2230	687	715	716
11:2330	569	578	573
12:0030	596	600	617
12:0230	643	670	676
12:0330	553	586	586
12:0630	781	810	807
12:0530	703	730	733
12:0630	708	747	748
12:0730	769	808	801
12:0830	715	743	745
12:0930	701	748	751

Table No.5

Computed Solids Circulation

CAFB Run 4

SOLIDS CIRCULATION, LB/HR

DAY:HOUR	CIRCULATING CAO	TOTAL SOLIDS TO REGEN	TOTAL SOLIDS FROM REGEN
12:1010	753	783	787
12:1030	850	879	886
12:1130	751	779	781
12:1830	773	801	806
12:1930	777	805	806
12:1630	598	675	681
12:1730	693	721	728
12:1810	715	743	748
12:1910	950	978	981
12:2010	711	949	740
12:2110	1028	1055	1061
12:2210	700	730	735
12:2310	900	930	933
13:0010	900	933	938
13:0110	911	939	945
13:0210	902	934	941
13:0310	914	943	947
13:0410	1001	1031	1034
13:0510	919	950	954
13:0610	918	948	953
13:0710	1153	1185	1193
13:0810	709	739	747
13:0910	713	743	749
13:1010	870	900	904
13:1110	750	779	784
13:1210	698	727	733
13:1310	694	724	732
13:1410	833	863	870
13:1510	833	863	870
13:1610	833	863	870
13:1710	833	863	870
13:1810	833	863	870
13:1910	833	863	870
13:2010	833	863	870
13:2110	1153	1183	1193
13:2210	713	743	749
13:2310	833	863	870
14:0010	833	863	870
14:0110	833	863	870
14:0210	833	863	870
14:0310	833	863	870
14:0410	833	863	870
14:0510	833	863	870
14:0610	1153	1183	1193

Table D-2

Computed Boiling Circulation

(AFB Run 1)

SOLIDS CIRCULATION, LB/HR

DAY:HOUR	CIRCULATION END	TOTAL TO BOILER	SOLIDS FROM BOILER
11:06:30	651	611	898
11:07:30	661	663	671
11:08:30			
11:09:30	654	710	715
11:10:30	646	674	676
11:11:30	639	644	645
11:12:30	645	691	691
11:13:30	701	653	661
11:14:30	704	711	613
11:15:30	716	760	769
11:16:30	676	651	661
11:17:30	719	765	760
11:18:30	693	635	631
11:19:30	716	765	765
11:20:30	654	670	670
20:20:30	1411	1501	1511
20:21:30	1411	1476	1453
20:22:30	1448	1495	1495
20:23:30	911	986	986
21:00:30	1353	1051	1303
21:01:30	930	998	1301
21:02:30	998	1002	1012
21:03:30	1000	1009	1030
21:04:30	971	995	999
21:05:30	908	930	935
21:06:30	917	963	966
21:07:30	935	992	997
21:08:30	961	990	990
21:09:30	951	975	977
21:10:30	941	975	979
21:11:30	975	995	997
21:12:30	990	951	960
21:13:30	935	955	955
21:14:30	915	945	945
21:15:30	919	965	965
21:16:30	938	980	980
21:17:30	971	995	995
21:18:30	1000	1000	1000
21:19:30	1000	1000	1000
21:20:30	1000	1000	1000

Table 8c8

Computed Bottom Circulation

APG Run 2

SOLIDS CIRCULATION: LB/HR

DAY:HOUR	CIRCULATING GAS	TOTAL TO BOTTOM	SOLIDS FROM BOTTOM
01:01:30	686	630	618
01:03:30	582	611	614
01:13:30			
02:00:30	607	630	611
02:01:30	611	636	630
02:03:30	630	706	730
02:10:30	598	660	661
02:13:30	553	606	604
02:16:30	644	670	631
02:21:30	607	630	630
02:18:30	636	700	706
02:06:30	585	653	655
03:00:30	651	630	653
03:13:30	657	691	670
03:11:30	619	651	651
03:17:30	577	606	579
03:13:30	616	645	601
03:14:30	591	676	681
03:15:30	666	633	653
03:16:30	656	650	630
03:17:30	607	653	633
03:19:30	634	676	661

Table No5

Computed Oxides Circulation

(APS Run 2)

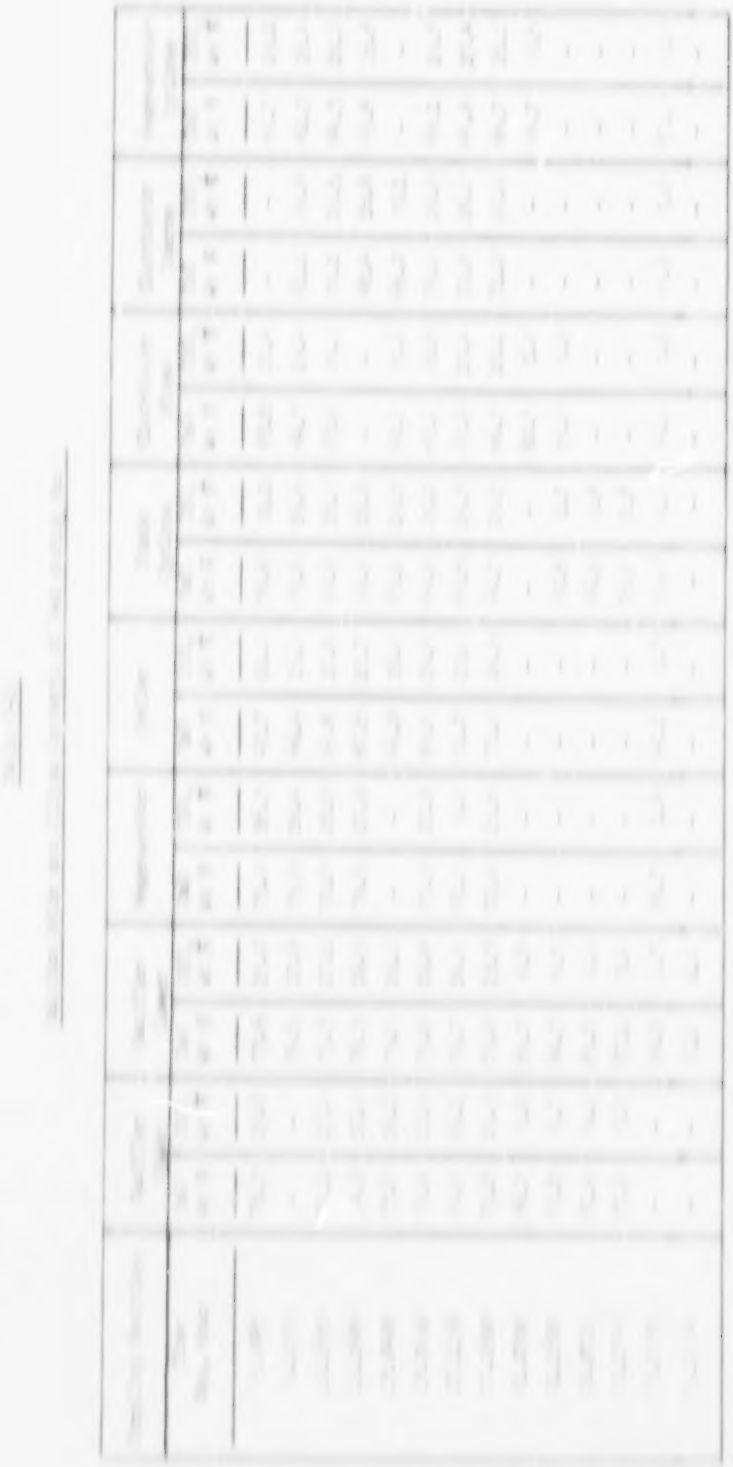
SOLIDS PRECIPITATION; ERROR

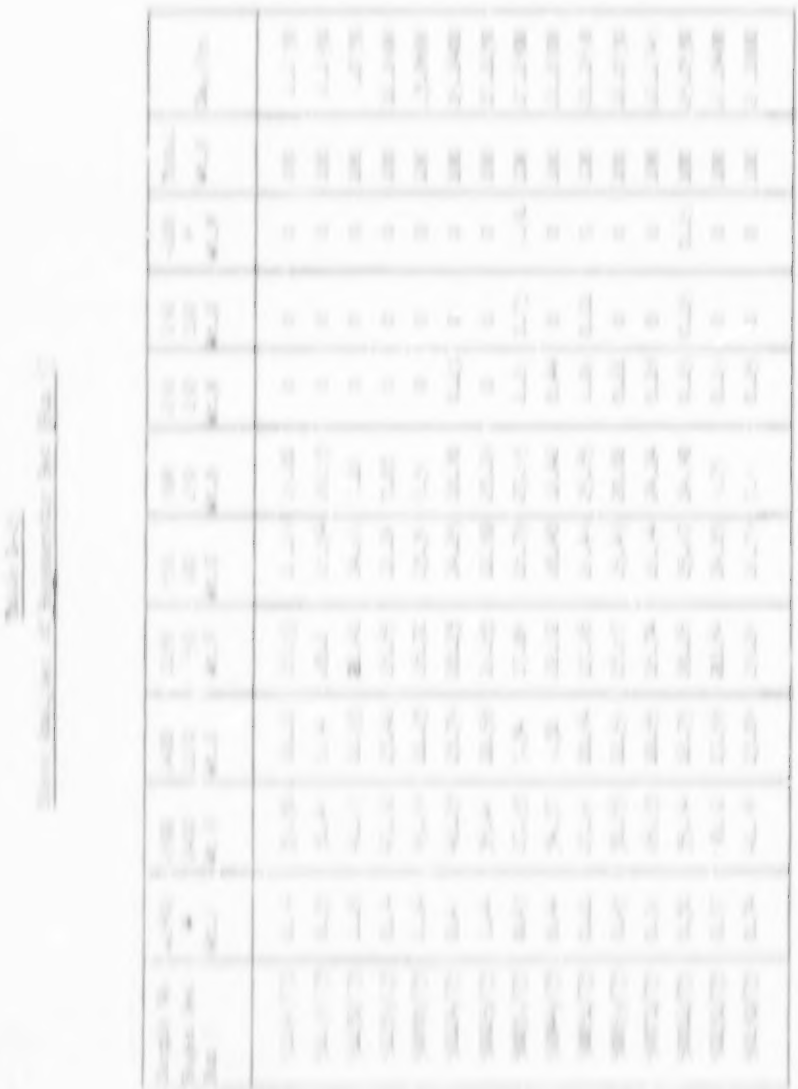
DAY:HOUR	PRECIPITATING END	TOTAL TO BEGIN	SOLIDS FROM BEGIN
03:17:30	319	966	368
03:18:30	358	990	389
03:19:30	377	1054	366
03:20:30	375	1050	811
03:21:30	663	1006	697
03:22:30	639	1016	677
03:23:30			
04:00:30	635	1078	615
04:01:30	709	1111	738
04:02:30	658	1079	586
04:03:30	691	1075	669
04:04:30	671	1066	701
04:05:30	659	1031	649
04:06:30	636	1051	710
04:07:30	618	1033	638
04:08:30	663	1059	689
04:09:30	716	1101	750
04:10:30	711	1135	769
04:11:30	733	1162	851
04:12:30	655	1121	851
04:13:30	670	1075	636
04:14:30	693	1110	707
04:15:30	683	1099	711
04:16:30	696	1100	713
04:17:30	669	1119	719
04:18:30	666	1097	715
04:19:30	670	1091	695
04:20:30	736	1164	733
04:21:30	771	1253	811
04:22:30	780	1193	777
04:23:30	830	1209	818
05:00:30	833	1231	883
05:01:30	915	1281	855
05:02:30	809	1206	819
05:03:30	755	1251	865
05:04:30	771	1210	824
05:05:30	775	1200	825
05:06:30	798	1251	811
05:07:30	796	1185	788
05:08:30	784	1211	838

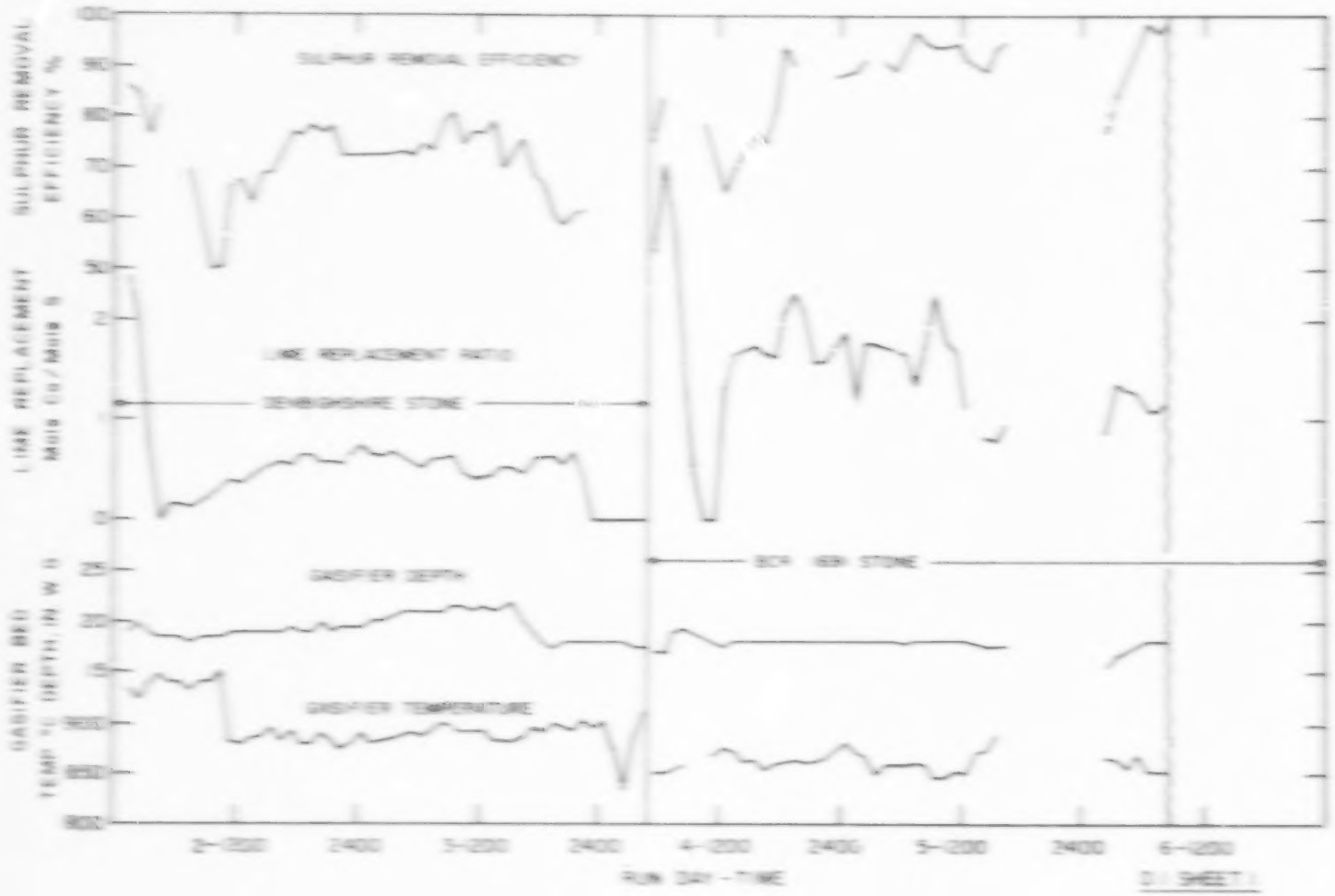
Table 2-1.

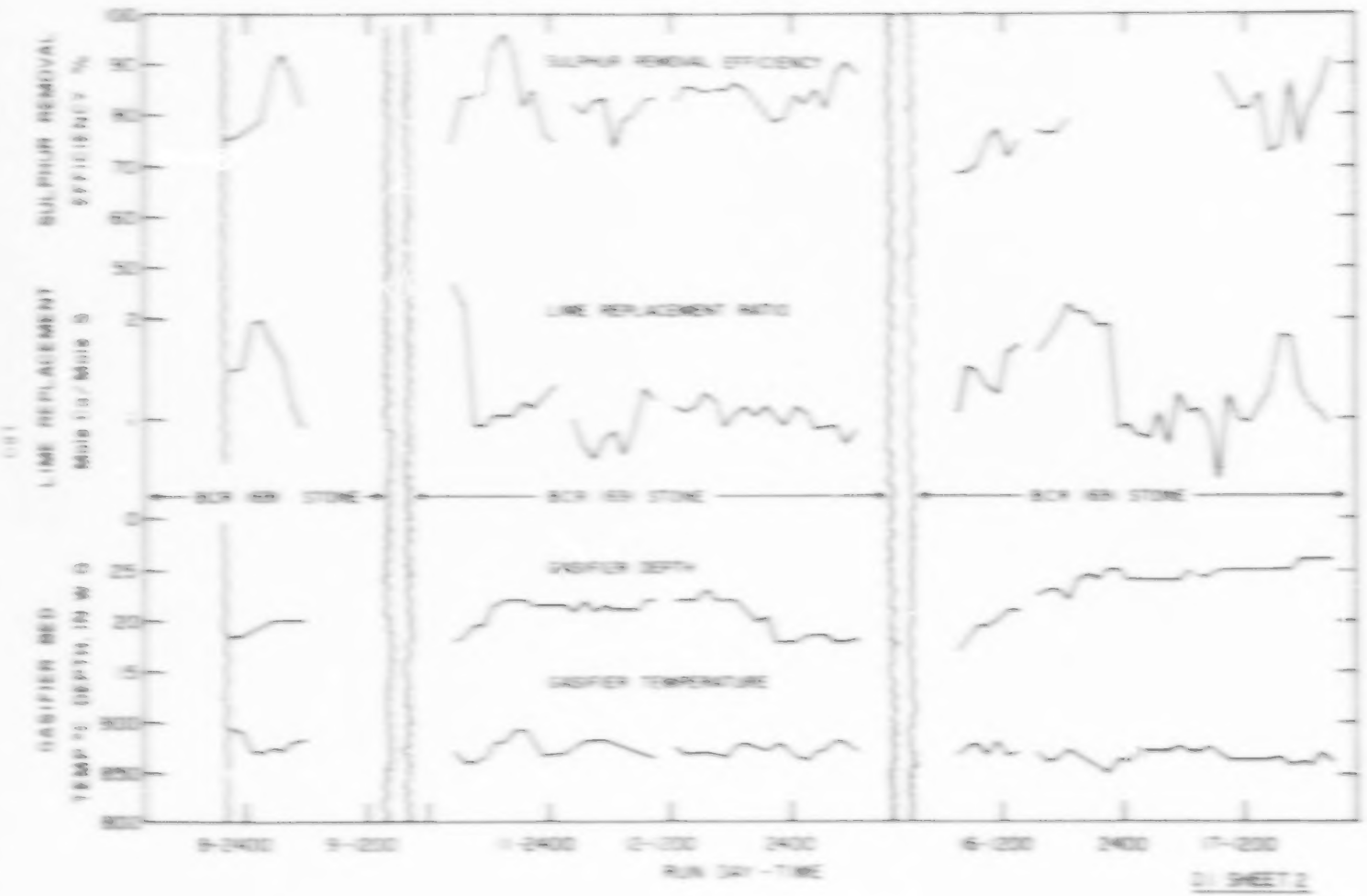
(LOR Reliable Control) (Rev. 1)

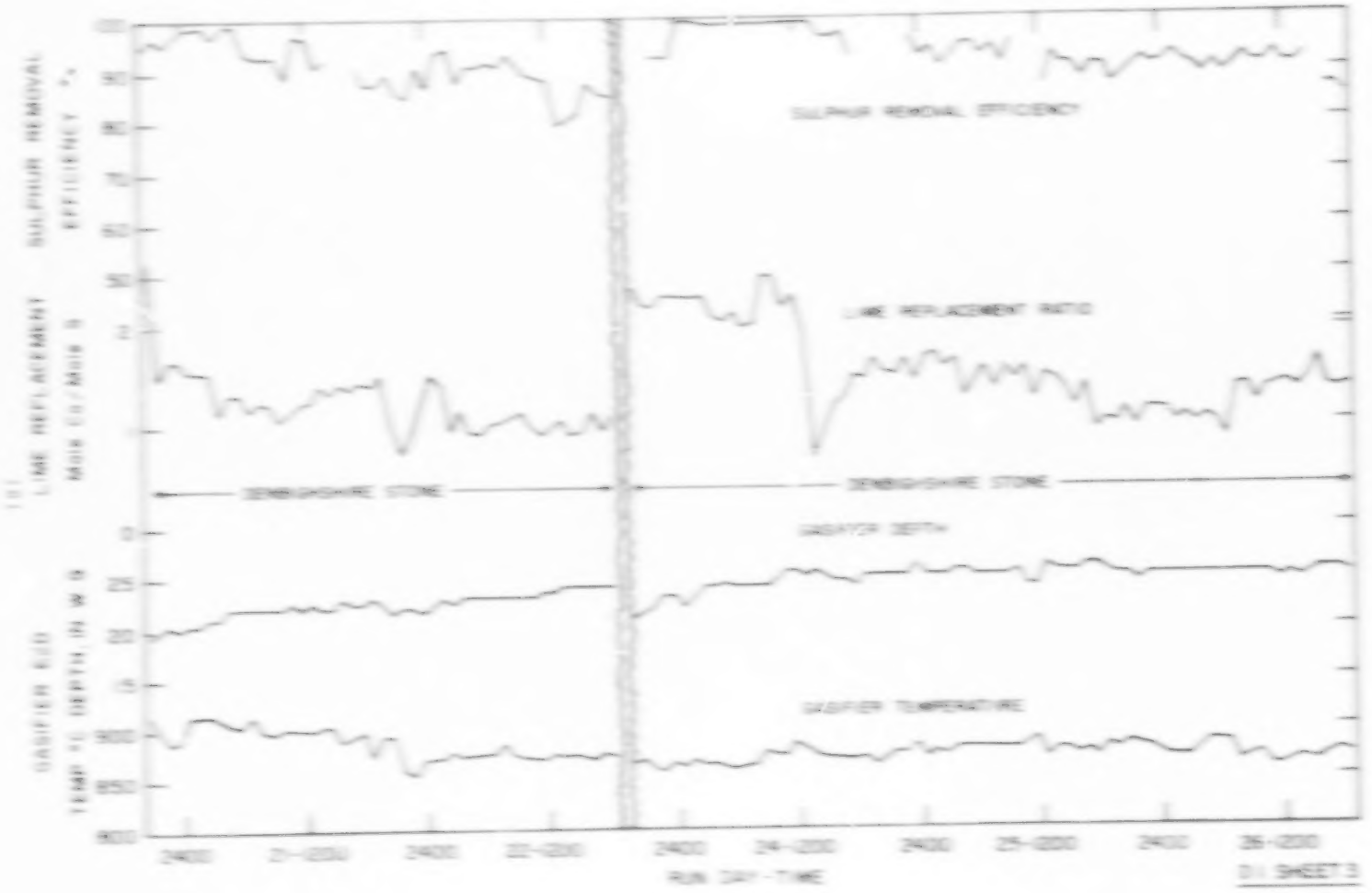
Time (Hr, Min)	Location Position	Variation (in)	Indication (in)	Signal (in)
7:15 A	Oscillator Lower	1000	750	877
	Oscillator Upper	1050	700	875
	Regenerator	1250	875	875
8:00	Oscillator Lower	1100	715	893
	Regenerator	1400	900	863
9:00	Oscillator Lower	1150	630	858
	Oscillator Upper	1300	720	856
	Regenerator	1600	1000	850
9:30	Oscillator Lower	1400	1100	900
	Regenerator	1600	1250	900
10:00	Oscillator Lower	1000	900	870
	Oscillator Upper	1200	900	900
	Regenerator	1400	1100	901
10:30	Oscillator Lower	1000	900	915
	Oscillator Upper	1200	917	937
	Regenerator	1400	917	901
11:00	Oscillator Lower	1100	900	900
	Oscillator Upper	1250	1000	900
	Regenerator	1400	1100	870
11:30	Oscillator Lower	1100	915	904
	Oscillator Upper	1200	920	904
	Regenerator	1500	1000	903
12:15 P	Oscillator Lower	1150	1100	925
	Oscillator Upper	1300	915	900
	Regenerator	1500	1000	900
12:30	Oscillator Lower	1000	915	900
	Oscillator Upper	1300	970	900
	Regenerator	1500	1000	900
1:00	Oscillator Lower	1000	900	955
	Oscillator Upper	1200	1000	958
	Regenerator	1500	1000	915
1:30	Oscillator Lower	1000	795	915
	Oscillator Upper	1300	1150	900
	Regenerator	1500	1000	900
2:00	Oscillator Lower	1100	900	900
	Oscillator Upper	1250	1000	900
	Regenerator	1500	1000	870
2:30	Oscillator Lower	1300	1000	885
	Oscillator Upper	1700	1300	870
	Regenerator	1900	1400	870
3:00	Oscillator Lower	1400	1000	900
	Oscillator Upper	1700	1300	870
	Regenerator	1900	1400	870
3:30	Oscillator Lower	1400	1000	900
	Oscillator Upper	1700	1300	870
	Regenerator	1900	1400	870
4:00	Oscillator Lower	1400	1000	900
	Oscillator Upper	1700	1300	870
	Regenerator	1900	1400	870
4:30	Oscillator Lower	1400	1000	900
	Oscillator Upper	1700	1300	870
	Regenerator	1900	1400	870
5:00	Oscillator Lower	1400	1000	900
	Oscillator Upper	1700	1300	870
	Regenerator	1900	1400	870



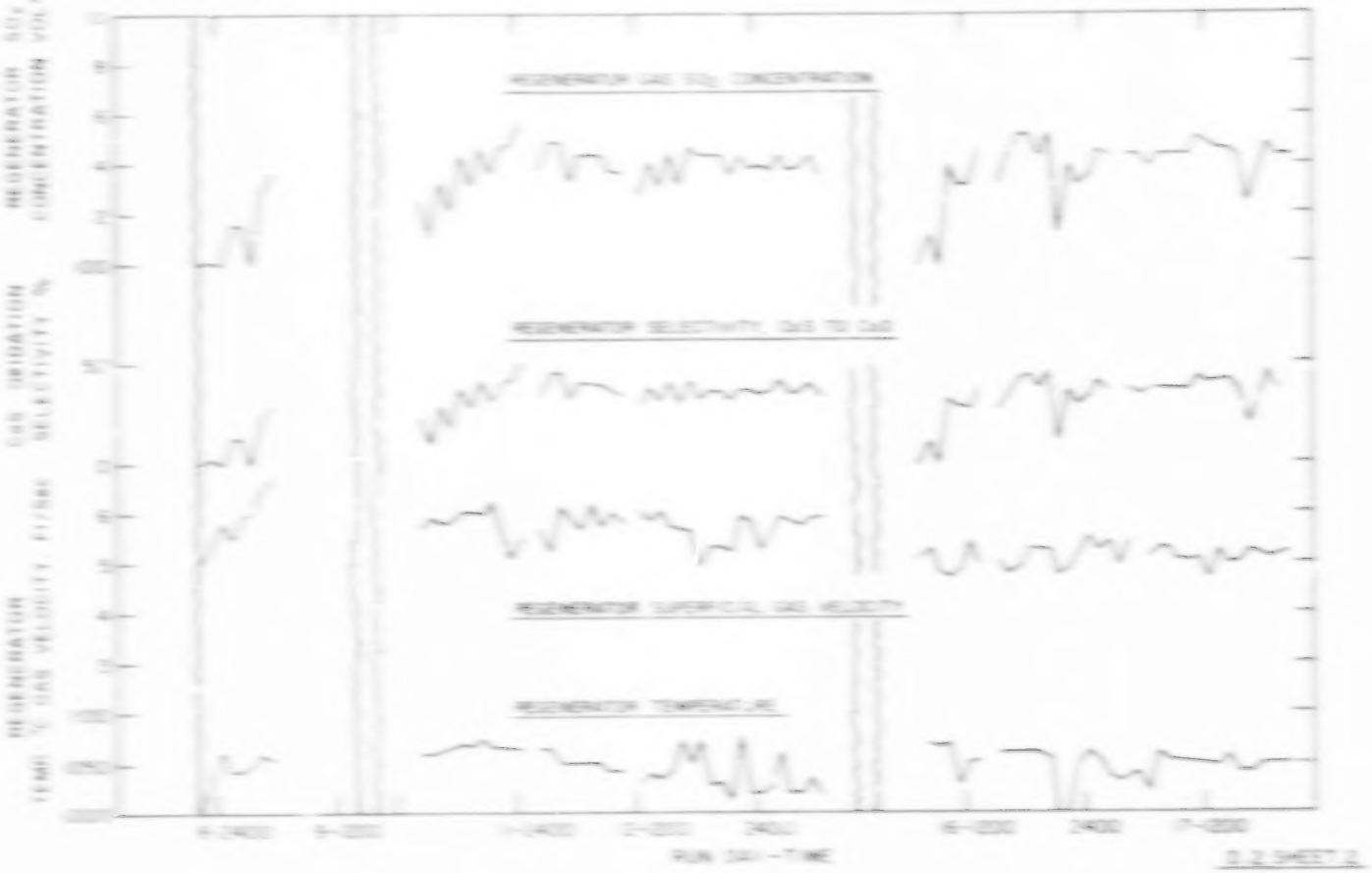


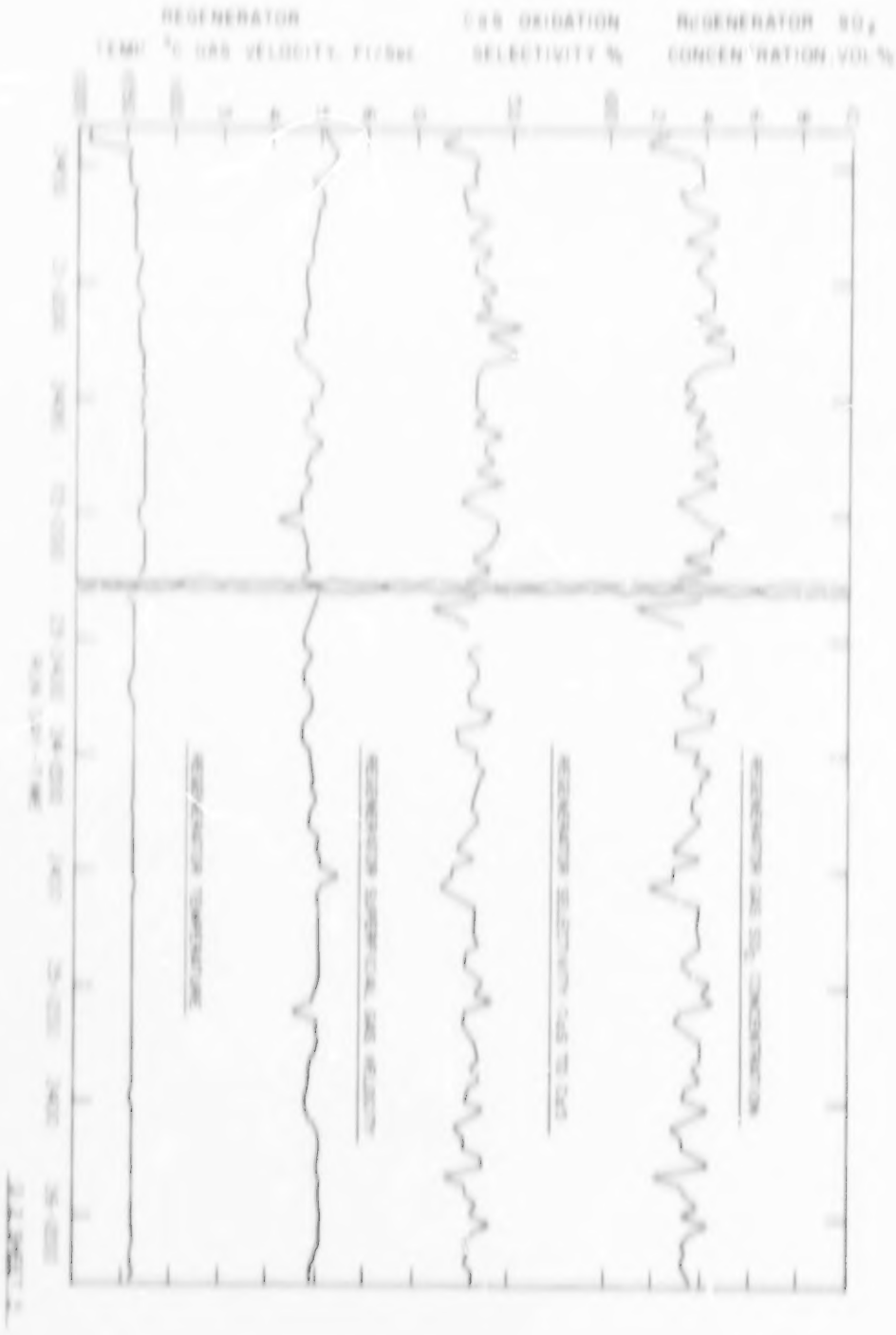

SULFUR REMOVAL EFFICIENCY
LIME REPLACEMENT RATIO
DENSIMORE STONE
BCH 459 STONE
DAMPER DEPTH
DAMPER TEMPERATURE
2-1200
2400
3-1200
2400
4-1200
2400
5-1200
2400
6-1200
Run Day - TIME


SULPHUR REMOVAL EFFICIENCY
LIME REPLACEMENT RATIO
BCR 459 STONE
BCR 459 STONE
BCR 459 STONE
9-2400
9-1200
1-2400
2-1200
2400
6-1200
2400
7-1200
RUN DAY - TIME



REGENERATOR
TEMP °F GAS VELOCITY FT/SEC
GAS ORIFICATION
SELECTIVITY %
REGENERATOR SO2
CONCENTRATION VOL %




REGENERATOR
TEMP °C
GAS VELOCITY, Ft/Sec
GAS OXIDATION
SELECTIVITY %
REGENERATOR SO2
CONCENTRATION VOL %

Appendix B

Gasifier Heat Balance

The heat released in the gasifier under partial oxidation conditions depends on the air fuel ratio and on the relative extents of the various competing reactions within the bed.

As a first step in development of a mathematical model to predict gasifier operating conditions, a series of heat and material balance equations have been programmed to compute the heat release rate in the gasifier from measured flow rates and temperatures. A second set of equations is used to estimate the heat release rate from rate of reaction and from estimates of the relative amounts of hydrogen and carbon oxidized and the CO/CO_2 ratio produced.

List of Variable Names

Variable	Meaning	Units
F	Fuel rate	lb/Hr
A	Air rate	SCFM
D	Gasifier bed depth	inches
S	Average bed particle size	microns
STNRF	Fresh stone feed rate	lb/Hr
TG	Gasifier bed temp	Deg.F
TR	Regenerator bed temp	Deg.F
V	Superficial gas velocity	ft/sec
CPCT	Carbon in fuel oil	wt.%
HPCT	Hydrogen in fuel oil	wt.%
SPCT	Sulphur in fuel oil	wt.%
CAOPCT	CaO in stone	wt.%
O2INPU	Oxygen in flue gas	Vol.%
STNCIR	Lime circulation rate	lb/Hr
Q	Flue gas inlet rate	SCFM
CO2PCT	Carbon dioxide in flue gas	Vol.%
TAIR	Temperature of inlet plenum	Deg.F
O2FED	Oxygen feed to gasifier	lb Mole/Hr
RATIO	CO/CO_2 ratio produced in gasifier bed combustion	Dimensionless
FRACO	Fraction of CO_2 in CO and CO_2 produced in gasifier bed combustion	Dimensionless
TG	Gasifier temperature	Deg.F
HCO2G	Heat of Combustion for C to CO_2	Btu/lb mole C
HCOG	Heat of combustion for C to CO	Btu/lb mole C
HH2OG	Heat of Combustion for H_2 to H_2O	Btu/lb mole H_2
STCO	Oxygen needed for stoichiometric combustion of fuel oil	lb Mole O_2/100 lb oil
R	Oxygen fed as percent of stoichiometric	percent
TRP	Regenerator temperature	Deg.F
HTXCIR	Heat transferred from regenerator to gasifier by circulating lime	Btu/Hr

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Variable	Meaning	Units
HTUOIL	Heat required to raise temperature of oil from inlet to gasifier temperature	Btu/Hr
HTU/HR	Heat required to crack oil to H_2 + C	Btu/Hr
HAIR	Heat required to raise inlet air from plenum temperature to 1000 Deg.F	Btu/Hr-GPM
HTUAIR	Heat required to raise inlet air from plenum temperature to gasifier temperature	Btu/Hr
C	Ratio of CaO to S fed	Dimensionless
HTU/AL	Heat required to calcine limestone	Btu/Hr
HTUCO2	Heat required to raise temperature of CO_2 from limestone to gasifier temperature	Btu/Hr
HTUSTN	Heat required to raise nonvolatile part of limestone to gasifier temperature	Btu/Hr
HTULOS	Heat lost from gasifier bed through walls	Btu/Hr
HTUFED	Net heat input to gasifier except for reaction and flue gas contributions	Btu/Hr
AA	Constant in equation for percent of carbon oxidized	Dimensionless
FOXDP	Portion of fuel carbon oxidized	Percent
COXIM	Amount of fuel carbon oxidized	lb Mole/100 lb fuel
HTUCO2	Heat released by oxidizing carbon to CO_2	Btu/100 lb fuel
HTUCO	Heat released by oxidizing carbon to CO	Btu/100 lb fuel
HTUH2O	Heat released by oxidizing hydrogen to H_2O	Btu/100 lb fuel
HTUFG	Heat required to raise flue gas from plenum to gasifier temperature	Btu/Hr
HTUBRN	Heat released by combustion in gasifier	Btu/Hr
HTUBRL	HTUBRN x 10^{-6}	(Btu/Hr) x 10^{-6}
HTUB	Heat released per lb oil by combustion in gasifier	Btu/lb oil
HOXD	Heat released per mole oxygen by combustion in gasifier	Btu/lb mole
HOXIM	Amount of fuel hydrogen oxidized in gasifier	lb Mole/100 lb fuel
HOXDP	Portion of fuel carbon oxidized	Percent
	HOXIM and HOXDP are based on heat balance calculation	
COXP	Amount of fuel carbon oxidized to CO_2	lb Mole/100 lb fuel
COM	Amount of fuel carbon oxidized to CO	lb Mole/100 lb fuel
O/H2O	Amount of oxygen converted to H_2O	lb Mole/100 lb fuel
HOXIMO	Amount of fuel hydrogen oxidized	lb Mole/100 lb fuel
HOXDPO	Portion of fuel hydrogen oxidized	Percent
	HOXIMO and HOXDPO are based on material balance calculation	
HOXFL	Heat release per mole oxygen by combustion in gasifier, calculated by stoichiometry and reaction heats	Btu/lb mole

Thermal Equations

The thermal equations are linearized forms which adjust for departure from tabulated values at 1000 deg.F.

$$H_{CO_2} = 10{,}730 + .436 * (T_0 - 1000)$$
$$H_{CO} = 7{,}805 + .275 * (T_0 - 1000)$$
$$H_{H_2O+O_2} = 10{,}951 + 1.357 * (T_0 - 1000)$$

The air heat equality is based on published enthalpies at 100°F and 1000°F and on specific heats (C_p) at these levels.

$$H_{AA} = \left(\frac{CO}{100} \frac{BTU}{LB\ MOLE} \frac{LB\ MOLE}{359.7\ CU.FT.}\right) * (H_{1,000} - (H_{100} + C_{p100}(TAIR - 100)))$$

H_{AA} = Enthalpy change to heat air from inlet temp. to 1000°F.
The term $\frac{CO}{100}$ converts from CUM to lb Mole/Hr

$$H_{AA} = .1579 * (10857 - (3805 + 6.90 * (TAIR - 1.0.)))$$

$$H_{TAIR} = A * (H_{AA} + (\frac{CO}{100} * C_{p1.00}) * (T_0 - 1000))$$

$$H_{TAIR} = A * (H_{AA} + .0439 * (T_0 - 1000))$$

The enthalpy change to heat the flue gas is based on a flue gas of assumed composition:

.448 O_2 + 75% N_2 + 10.1% H_2O + 12.5% CO_2

Weighted average values of enthalpies and specific heats were combined to obtain the final equation.

$$H_{TURB} = 3 * (\frac{CO}{100}) * (H_{1000} - H_{300} + C_{p1000} * (T_0 - 1000) + C_{p300}(300 - TAIR))$$

$$H_{TURB} = 3 * .1579 * (10719 + 6.90 * T_0 - 1845 + 7.403 * TAIR + .207)$$

$$H_{TURB} = 3 * (1,4016 * T_0 + 1.172 * TAIR + 204.79)$$

Heat of limestone calcination is taken at 1910 Btu/lb of CaO

$$H_{RVCAL} = H_{TURB} * CAOUT * .01 * 1910$$
$$H_{RVCAL} = H_{TURB} * CAOUT * 19.10$$

Heat to raise temperature of CO_2 liberated in calcination:

$$HTCAO_2 = START * CAORXT * \frac{.01}{44} * (H_{1200} + C_{p1200}(TG-1200))$$

$$HTCAO_2 = START * CAORXT * \frac{.01}{44} * (17935 + 13.36\ (TG-1200))$$

$$HTCAO_2 = START * CAORXT * (3.03 + .00639 * (TG-1200))$$

Heat to raise temperature of nonvolatile portion of stone:

$$HTSTN = (START) * (Fraction\ Nonvolatile) * (H_{1200} + C_{p1200} * (TG-1200))$$

$$HTSTN = START * (1 - \frac{11}{50} * \frac{CAORXT}{100} * (362.9 + .285 * (TG-1200))$$

$$HTSTN = START * (1 - .00786 * CAORXT) * (362.9 + .285 * (TG-1200))$$

Heat supplied to oil:

Heat uptake by the oil is made up of sensible heat, latent heat, and heat of cracking.

Sensible heat and latent heat are included in a linear equation based on the oil enthalpy at 1200 °F and the specific heat of the oil between 1200°F and gasifier temperature.

$$HTOIL = P * (780. + 0.54 * (TG-1200))$$

The heat of cracking is assumed to be 600 Btu/lb

$$HTCRK = P * 600$$

The heat loss from the gasifier bed is estimated from an analysis which indicated that the effective product of thermal conductivity and bed area is equal to 0.282 x D. The heat loss equation is therefore:

$$HTLOS = 0.282 * D * (TG-70)$$

Material Balance Equations

Oxygen fed:

$$O2FED = 0.21 * \frac{32}{359} * A + \frac{32}{359} * Q + \frac{O2INPT}{170}$$

$$O2FED = .0130 * A + .00158 * Q + O2INPT$$

CO/CO_2 Ratio

The ratio of CO to CO_2 produced by partial combustion of oil in the fluid bed is a complex function of thermodynamics, reaction kinetics and contacting in the bed.

Analytical results indicate that the values obtained are much lower than equilibrium for the temperatures encountered.

It is believed that considerable CO_2 forms by reaction of carbon on lime in the highly oxidizing region near the air inlet nozzles and that insufficient contacting time is available for this CO_2 to reach equilibrium with CO and carbon in the upper portion of the bed.

The equation used here is an empirical equation relating CO/CO_2 ratio to temperature, based on gas analysis data obtained in the batch CFBR reactors during the phase I study and reported in Appendix D of reference (1). The equation is:

$$RATIO = 2.91\ E{+}4 * EXP(9.76\ E{+}3 * RG)$$

Fraction of Carbon Oxidized

In the current analysis it is assumed that components of the flue gas, recycled for temperature control, do not react in the gasifier. The validity of this assumption remains to be verified. With this assumption, fixing the CO/CO_2 ratio and the fraction of feed carbon oxidized also fixes the fraction of feed hydrogen oxidized since the oxygen fed must appear as CO, CO_2 or H_2O.

The fraction of carbon oxidized is based on an empirical equation first derived from batch unit studies but modified to improve match with the heat results of the continuous pilot plant runs. This equation is:

$$COXDF = EXP(-AA) * (R\,\Phi.25) * (RO\,\Phi1.33)$$

A value of 0.4 for (AA) has given the best match between measured and calculated heat release rates.

The moles of carbon oxidized per 100 lb of fuel is the product of the fuel carbon content and fraction oxidized.

$$COXDM = \frac{1}{12} * CHFF * \frac{COXDF}{100}$$

$$COXDM = 8.333\ E{-}4 * CHFF * COXDF$$

Calculated Heat Release of Fuel Oil

The calculated heat release per 100 pounds of fuel is given, for each element, by the moles of that element oxidized multiplied by the heat of combustion of that element.

$$HR_{CO_2C} = MOC_2 * COXIM * PMC_2$$

$$HR_{COC} = MOC * COXIM * (1 - PMC_2)$$

$$HR_{H_2O} = MH_2O * MOXIM$$

The total heat release is the sum of that for the elements.

"Measured" Heat Release

The "measured" heat release is the difference between the sensible heat inputs to the gasifier and the heat required to raise the products to gasifier temperature.

$$HR_{GAS} = HR_{AIR} + HR_{ASH} + HR_{CAL} + HR_{OIL} + HR_{MOIS} + HR_{AIR} + HR_{STM} + HR_{STM} + HR_{AIR}$$

Hydrogen Oxidized

The amount of hydrogen oxidized can be estimated from the heat balance or from the oxygen balance.

Derivation of the heat balance equation for MOXIM is as follows:

$$\text{Heat release per mole } O_2 = \frac{\text{Heat of Combustion of elements}}{O_2 \text{ needed to burn elements}}$$

$$MOXD = \frac{HR_{CO_2C} + HR_{COC} + HR_{H_2O}}{(.5 * (MOXIM + COXIM * (1 - PMC_2)))}$$

Also, from the measured heat release in the gasifier,

$$MOXD = HR_{GAS} / PSI$$

Substituting, $HR_{H_2O} = MOXIM * MOC_2C$

gives,

$$MOXD = \frac{HR_{CO_2C} + HR_{COC} + MOC_2C * MOXIM}{(.5 * (MOXIM + COXIM * (1 - PMC_2)))}$$

Rearranging and solving for [illegible] gives:

[illegible]

Material balance hydrogen satisfied

Since all oxygen fed to the gasifier is assumed to exist as O_2, CO_2, or H_2O, fixing the amounts of CO and CO_2 by empirical expressions also fixes the quantity of H_2O.

[illegible]

Calculated Heat Release (y, Btu/lb O)

A purely calculated heat release per lb oxygen can be computed for comparison with the "measured" value. For this purpose a value of $HR_{BTU/LB}$ is calculated using the material balance value of hydrogen satisfied.

$$HR_{BTU/LB} = HOXIDE \times HR_{BTU/LB\text{-}O}$$

then,

$$HREL = .01 \times F \times (HR_{COCO_2} + HR_{STEAM} + HR_{BTU/LB\text{-}O})/O_{FED}$$

These heat and material balance equations have been used in several computer programs to analyse experimental data and to predict new operating conditions. Table B-1 list the FORTRAN statements of program HEATBC which calculates the gasifier heat release from experimental operating conditions and compares the value with the predicted value. Results of calculations on CRPU data appear in Table B-2.

Table B-1. Fortran Listing of
CAPE Heat Balance Programs

JHTHOC

```
1000 CALCULATE HEAT OF COMBUSTION FROM CARB CONDITIONS
1100   PROGRAM NAME JHTHOC
1200   PRINT 600
1300   DIMENSION P(25)
1400 $FILE JHTF(1)
1500 50 CONTINUE
1600 READ (1) (P(I), I=1,17)
1700 IF (ENDFILE 1) 10, 600
1800 10 CONTINUE
1900C DEFINE VARIABLE VALUES
2000   100 F = P(1)     'FUEL RATE, LB/HR
2100   A = P(2)
2200 D = P(3)
2300   S = P(4)         'AVERAGE PARTICLE SIZE, MICRONS
2400   SINRT = P(5)
2500   TC = P(6)        'GASIFIER TEMP DEG C
2600   TR = P(7)        'REGENERATOR TEMP DEG C
2700   V = P(8)         'GAS VELOCITY (INITIAL ASSUMPTION)
2800   CPCT = P(9)      'CARBON IN FUEL, WT. PCT
2900   HPCT = P(10)     'HYDROGEN IN FUEL, WT. PCT
3000   SPCT = P(11)     'SULFUR IN FUEL, WT. PCT
3100   CAOPCT = P(12)   'CAO IN STONE, WT. PCT
3200   OPINFG = P(13)   'OP IN FLUE GAS, VOL PCT
3300   SINCIR = P(14)   'LIME CIRCULATION RATE, LB/HR
3400 G = P(15)
3500   CODEG = P(16)
3600   TAIN = P(17)*(1.8)+32.         'AIR INLET TEMP, DEG F.
3700   TO   = 1.8*TC  + 32.
3800   TRF  = 1.8*TR+32.
3900   BTUCIR = SINCIR*(TRF-TO)*.28A   'HEAT SUPPLIED BY CIRCULATING LIME
4000   BTUOIL = F*(778. + .8*(TC-1000.))
4100   BTUCRK = F*480.                          ' HEAT TO CRACK OIL
4200   BHA = .15794*(15857.+(3825.+6.8*(TAIN-100.)))
4300   BTUAIR = A*(BHA+.8*39*(TC-1600.))
4400 C = STNRT*CAOPCT/(1.756*SPCT)
4500 BTUCAL = STNRT*CAOPCT*(A-1
4600 BTUCOP = STNRT*CAOPCT*(3.083+.0023*(TC-1600.))
4700   BTUSTN = STNRT*(1.+.88*BHA*CAOPCT)*(360.9+.28*(TC-1600.))
4800 BTULOS = .04*(TC-70.)
4900   BTUFXD = BTUCIR+BTULOS+BTUCAL+BTUOIL+BTUCRK+BTUAIR+BTUSTN+BTUCOP
5000   BTUFG = (1.485*TC-1.17*TAIN+284.79)*G
5100   BTURBN = BTUFG-BTUFXD
5200 HTREL = BTURBN/(1.E+6)
5300   BTULB = BTURBN/F
5400   COFED = .833*BHA+.88*SR*CO*OPINFG    'LBLES OF TO GASIFIER/HR
5500   HOXD = BTURBN/COFED
5600   ISR RATIO = 0.91E-4*EXP(9.7E-3*TC)
5700   FRCOP = 1./(RATIO + 1.)              'FRACTION COP IN PRODUCT GAS
5800 STOP = CPCT/12. +HPCT/4. + SPCT/32.    'STOIC OP, MOL/100 LB
5900 R = 1.E+6*COFED/(F*STOP)
```

Table B-1 Fortran Listing of
APH Heat Balance Programs

```
(WITHOUT CONTINUED

600 AA = 8.4
610 200 COXOP = EXP(-AA(1)(1).84219(TC)(.1314)
620  COXOM = 8.1118*AA*CPCT*COXOP   'MOLE C OXIDIZED/100 LB FUEL
630  HCCOP = 169790. + .4349(TC - 1460.)
640  HCCO = 48525 + .7375*(TC-1460.)   ' REACT HEAT, C TO CO
650  HHPHPO = 104794. + 1.327*(TC-1460.)   '  REACT HEAT,  H2 TO H2O
660  BTUCCOP = HCCO*COXOM*FRCOP     'BTU/100 LB FUEL FOR C TO CO
670 BTUCCO = HCCO*COXOM*(1.-FRCOP)
680 HOXOM=(BTUCCOP-BTUCCO-(1.5*COXOM*(1.-FRCOP))*HO2BO)/(1.5*HO2BO-HHPHPO)
690 HOXOP = 2.*P*HOXOM/HPCT
700 COPM = COXOM*FRCOP
710 COM = COXOM-COPM
720 OPHPO = OPFEO*100./F -COPM-COM/2.
730 HOXOM2 = OPHPO*P.
740 HOXOP2 = HOXOM/(HPCT*5.E-3)
750 BTUHPO = HOXOM*P*HHPHPO
760 HOEL = .01*F*(BTUCCOP+BTUCCO+BTUHPO)/OPFEO
765 PRINT ,HOXOP,HOXOP2
780 610 FORMAT (4X,F5.1,4X,F6.2,4X,F7.0,4X,F8.0,4X,F7.0,4X,F5.1)
790 GO TO 50
800 600 FORMAT (/,6X,3HB,7X,5HHTREL,4X,6HBTU/LB,6X,6HBTU/BP)
810 800 STOP
820  END
```

Appendix F

Gasifier Product Composition

Samples of gasifier product were collected in metal containers and analyzed for composition several times during run 5. Since the coke, tar, and heavy hydrocarbon portion of the product is not measured by this procedure, an attempt has been made to estimate the quantity and composition by material balance on the gasifier itself.

The dry gas compositions measured were listed in Table 19 of the text, and a summary of the material balance results and estimated heavy product compositions was listed in Table 20.

The calculations made involve several assumptions. Examples of these calculations and assumptions are presented in this appendix.

Data from the operating period 05:0130 are used in the examples.

- Air Flow to Gasifier, SCFM

23	To plasma and fuel injectors	
2.8	with Diene feed	
?	to cyclone inlet	
25.8	SCFM	

O_2 Feed = 25.8 x .0510 $\frac{\text{lb Mole } O_2 / \text{Hr}}{\text{SCFM Air}}$

= 1.313 lb Mole/Hr O_2

- Flue Gas to Gasifier

Oxygen and CO_2 are measured on dried samples of gas in continuous analyzers. The rate of flue gas flow to the gasifier is measured by orifice meter on wet gas. A correction is therefore needed for composition change due to water.

$$\text{Dry Gas Rate} = (\text{Wet Gas Rate}) \left(\frac{100-Z}{100-Y+.17Y} \right)$$

Where Y = Vol.% CO_2 in Dry Gas
X = Vol.% H_2O in Wet Gas
Z = Vol.% CO_2 in Wet Gas

The ratio X/Z is assumed to be .810 based on the composition of the fuel oil. This assumption implies that no portion of the flue gas does not change during passage through the flue gas scrubber.

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For the sample period, [illegible]

Wet Gas Rate = [illegible] SCFM

Dry Gas Rate = [illegible] = [illegible] SCFM

Dry Gas Rate = [illegible] = [illegible] lb Mole/hr

<u>Inputs with Flue Gas</u>

Mole/Hr O_2 = Mole/Hr Flue Gas x Mole Fraction O_2

CO_2 = [illegible] Mole/hr
O_2 = [illegible]
N_2 = [illegible]

Mole/Hr H_2O = Wet Gas Rate - Dry Gas Rate

H_2O = [illegible] = [illegible]

<u>Inputs from Solids</u>

<u>Oxygen from [illegible] Sulphate</u>

[illegible]

O_2 from Sulphate [illegible] Mole/Hr

<u>Oxygen from [illegible]</u>

[illegible]

[illegible] Mole O_2/Hr

<u>Oxygen from CaO + H_2O → [illegible] + H_2O</u>

[illegible]

[illegible] Mole O_2/Hr

<u>Total O_2 Inputs to Facility</u>

Air	[illegible]
Flue Gas O_2	[illegible]
CO_2	[illegible]
H_2O	[illegible]
Sulphate	[illegible]
Stone CO_2	[illegible]
CaO	[illegible]
	11,[illegible] lb Mole O_2/Hr

Nitrogen Inputs

Gas Air + Cyclone Inlet Flow - Stone Feed Air

= O_2 Rate x .79/.21

= 7,819 x .79/.21 = 29,415 Mol/Hr N_2

Gas Feed N_2 Bleed	= .361	Mole/Hr
Injector N_2	= .16	Mole/Hr
Flue Gas	= 10.19	Mole/Hr
Pressure Tapping Bleeds	= .03	
Total N_2 Inputs	30.185	Mole/Hr

Product Gas Rate = N_2 Rate/fraction N_2

= 30.185/.635 = 63.321 Mole/Hr

Oxygen Outputs in Product Gas 2 oxides

= (Fraction CO_2 + FrCO/2) (Product Gas Rate)

= (.1000 + .0930/2) (63.321) = 9.05 Mole O_2 /Hr

O_2 Out as Sulphate

= (Lime Circulation) (Fr S as SO_4^{--}/32) x 2

= 5/5 x .0829/16 = .149 Mole O_2/Hr

O_2 Out as H_2O

The oxygen leaving the gasifier as water is assumed to be the difference between the O_2 inputs and other known O_2 outputs.

O_2 in Water = 11.079 - 9.05 - .149 = 1.68 Mole/Hr

H_2 Inputs to Gasifier

Hydrogen in with Fuel = Fuel Rate x wt. fraction H_2/2

= 909.1 x 0.114/2 = 51.319 Mole H_2 /Hr

Hydrogen in with Flue Gas = 1.909

Total H_2 In = 59.703 Mole/Hr

Hydrogen Output in Product H_2 + Hydrocarbons

$\sum$ (Mole/Hr Product) (Mole fraction) x H_2 Multiple)

As H_2 = (3.5) x .(65%) = 1,050
As CH_4 = (3.5) x .(049) x 2 = 7,716
As C_2H_6 = (3.5) x .(03) x 3 = 4,725

Total 13,491

Hydrogen Out as H_2O

Moles H_2 in Water = 2 x Moles O_2 in water

H_2 out in Water = 2 x 1,68 = 3,356

Hydrogen Missing

= Known Inputs - Known Outputs

= (4,773 + 10,000 + 3,356 = 4,765 Mole H_2/Hr

Missing hydrogen is assumed to be in tars, coke and liquid portions of product not measured by gas chromatograph.

Carbon Inputs

With Fuel = Fuel Rate x fraction C/lb
= 909.1 x .056/1/ = 29,182 Mole C/Hr

With flue gas = CO_2 input = 1,717 Mole C/Hr
With limestone = Stone Rate x Mole fraction CO_2

= 3/ x .439/44 = .316 Mole C/Hr

Total Carbon In = 31,215 Mole/Hr

Carbon Outputs

Carbon Out in Product Gas

= Σ Product gas x Mole fraction x # multiple

As CO_2 = (3.5) x .1860 = 6.084
As CO = (3.5) x .0710 = 5.247
As CH_4 = (3.5) x .0460 = 3.860
As C_2H_4= (3.5) x .0301 x 2 = 5.005

Dry gas total C = (1.937 lb-mole/Hr

Carbon Out Regenerator

= (Mole/Hr Regenerator Gas) x (Mole fraction CO)

= N. to Regenerator = 1.00 Mole/Hr
N. to regenerator Gas = 0.1 Mole fraction
CO_2 in Regenerator Gas = .017 Mole fraction

CO_2 out = 1.00 x $\frac{.017}{.167}$ = .051 Mol./Hr

Carbon Missing

= Known Inputs - Known Outputs

= 3.015 - (1.937 + .051) = 3.992 Mole C/Hr

like hydrogen, the missing carbon is assumed to be in the form of C_2s, tar, and other heavy components of the product gas not measured by the gas chromatograph.

CO_2 Make in Gasifier

The CO_2 make in the gasifier is taken as CO_2 out less CO_2 from lime stone and flue gas recycle

CO_2 make = 6.809 - .316 - 1.717 = 4.856

CO make = 5.247

CO/CO_2 = 5.247/4.856 = 1.1

H/C Ratio of Heavy Products

The H/C ratio of missing products is taken as 2 x H_2 missing/C missing

H/C = 2 x 4.715/3.992 = .99

Results of these component material balance calculations for four gas samples are summarized here in Table F-1 and in Table 26 of the text.

Table F-1

Summary of Gasifier Component Material Balances

Flow	22:1030	22:1730	26:0130	26:1730
O_2 In, Mole/Hr	11.797	11.71[illegible]	11.679	11.[illegible]80
O_2 Out,	10.0[illegible]3	9.753	9.[illegible]99	8.951
O_2 to H_2O (Diff)	1.713	1.9[illegible]0	1.678	2.[illegible]09
H_2 to H_2O	5.4[illegible]6	5.[illegible]4	5.[illegible]0	5.618
H_2 Inputs	24.[illegible]01	24.[illegible]0	24.723	24.677
H_2 in products	20.97[illegible]	19.7[illegible]6	16.[illegible]6	12.301
H_2 missing	1.[illegible]03	1.[illegible]30	5.[illegible]66	7.758
C Inputs	31.[illegible]91	31.[illegible]06	31.[illegible]15	31.[illegible]56
C in Products	29.[illegible]34	29.[illegible]99	21.[illegible]32	20.[illegible]07
C Missing	2.[illegible]57	1.[illegible]07	9.[illegible]83	10.[illegible]49
H_2O in missing	.[illegible]86	.[illegible]58	.[illegible]06	1.93
CO/CO_2 Mode	1.[illegible]0	.[illegible]7	1.[illegible]0	.[illegible]6
H missing, % of feed	5.3	4.6	[illegible]0.4	35.5
H oxidized % of feed	19.[illegible]	19.[illegible]	19.9	19.0
C missing, % of feed	[illegible]5.9	3[illegible].0	33.1	37.5
C oxidized, % of feed	35.0	34.0	37.0	31.9